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REFINERY CATALYTIC CRACKER REGENERATOR SO_x CONTROL - STEAM STRIPPER LABORATORY TEST

T. Ctvrtnicek, et al

Monsanto Research Corporation

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ABSTRACT

The report summarizes experimental results from steam contacting of spent catalyst used in petroleum refinery fluid catalytic crackers. This concept has been identified as a potentially effective means of sulfur emission control for fluid catalytic cracker regenerators. Correlations between sulfur removal efficiency from the catalyst and the product of steam residence time in the stripper with the steam stripping rate are presented for several stripper designs. The extent of by-product formation, a discussion of pertinent equipment design, and recommendations for further investigation and development of this concept are also included. Additionally, the economics are presented as a function of steam stripping rate and fluid catalytic cracker unit size.

TABLE OF CONTENTS

Page

1. CONCLUSIONS 1
2. RECOMMENDATIONS 8
3. INTRODUCTION 10
4. EXPERIMENTAL WORK 12
  4.1 EXPERIMENTAL EQUIPMENT 12
    4.1.1 Catalyst Test Unit 12
    4.1.2 Spent Catalyst Procurement 14
    4.1.3 Catalyst Handling 19
    4.1.4 Steam Stripping Experiments and Catalyst Characterization 22
      4.1.4.1 Catalyst Steam Stripping 22
      4.1.4.2 Determination of the Sulfur Content of Coke 25
      4.1.4.3 Determination of the H_2O Content of Spent FCC Catalyst 25
      4.1.4.4 Determination of the Coke Content of Spent FCC Catalyst 26
    4.1.5 Other Experiments 26
    4.1.6 Analysis 27
      4.1.6.1 Sulfur Compound and Hydrocarbon Determination 27
      4.1.6.2 Analysis for the Total Organic Carbon Content of Stripper Condensate 28
      4.1.6.3 Volatility of Coke on Spent Catalyst 28
  4.2 EXPERIMENTAL RESULTS 28
    4.2.1 Catalyst Characterization 28
    4.2.2 Catalyst Steam Stripping 31
    4.2.3 Hydrocarbon Volatilization During Steam Stripping 33
    4.2.4 Volatility of Coke on Spent Catalyst 69
    4.2.5 By-Product Formation During Steam Stripping 73
    4.2.6 Effect of Steam Stripping on Catalyst Activity 75

TABLE OF CONTENTS (Continued)

Page

4.3 DISCUSSION OF RESULTS 81
4.4 DATA REGRESSION ANALYSIS 88
5. STEAM STRIPPING PROCESS DESIGN 102
5.1 CATALYST STEAM STRIPPER DESIGN 102
5.1.1 Semi-Batch Fluidized Bed Reactor 103
5.1.2 Rate Controlling Factors 110
5.1.3 Other Stripper Designs 112
5.1.3.1 Continuous Fluidized Bed Reactor 113
5.1.3.2 Plug-Flow Stripper 116
5.1.3.3 Counter-Current Stagewise Contacting 120
5.2 CONDENSER DESIGN 124
5.3 ACIDIFIER/PHASE SEPARATOR DESIGN 126
5.3.1 Equilibrium Relationship 126
5.3.2 Combined Condenser/Acidifier/Phase Separator Design (Alternative Sour Water Treatment System) 138
6. ECONOMICS 141
APPENDIX A SPENT FCC CATALYST STEAM STRIPPING DATA SHEET 154
APPENDIX B DETAILED COST ESTIMATES OF THE STEAM STRIPPING PROCESS 164

LIST OF FIGURES

Figure		Page
1	Control panel, reactor, and analysis system	13
2	Catalyst test unit, schematic flow diagram	15
3	Motionless mixer used in the catalyst reactor	16
4	Water deaeration system	17
5	Photograph of catalyst storage container with charging flask attached	21
6	Weighing flask mounted on catalyst charging bomb	23
7	Charging of the catalyst into a hot reactor	24
8	Results of steam stripping experiments on B-Series catalyst	35
9	Results of steam stripping experiments on C-Series catalyst	38
10	Results of steam stripping experiments on D-Series catalyst	40
11	Results of steam stripping experiments on E-Series catalyst with motionless mixer in stripper	43
12	Results of steam stripping experiments on E-Series catalyst without motionless mixer used in the stripper	45
13	Results of steam stripping experiments on F-Series catalyst	49
14	Results of steam stripping experiments on H-Series catalyst	53
15	Results of steam stripping experiments of H-Series catalyst	55

LIST OF FIGURES (Continued)

Figure		Page
16	Results of steam stripping experiments on I-Series catalyst	58
17	Effect of steam stripping rate upon total organic carbon content of stripper condensate	70
18	Summary of results of the regression analysis performed on B-Series catalyst	93
19	Summary of results of the regression analysis performed on C-Series catalyst	94
20	Summary of results of the regression analysis performed on D-Series catalyst	95
21	Summary of results of the regression analysis performed on E-Series catalyst	96
22	Summary of results of the regression analysis performed on F-Series catalyst	97
23	Summary of results of the regression analysis performed on H-Series catalyst	98
24	Summary of results of the regression analysis performed on H-Series catalyst	99
25	Summary of results of the regression analysis performed on I-Series catalyst	100
26	Schematic of spent catalyst steam stripper	104
27	Schematic of continuous fluidized bed stripper	114
28	Schematic of plug flow steam stripper	117
29	Counter-current, stagewise contactor	121
30	Distribution diagram for hydrogen sulfide (Note that $[S^=]$ has appreciable concentration only in strongly basic solutions	130
31	Effect of temperature upon ionization constant for H_2S	132

LIST OF FIGURES (Continued)

Figure		Page
32	H_2S concentration in vapor and liquid stream	136
33	Combined condenser, acidifier, phase separator clarifier, and scum oil removal system	139
34	FCC catalyst steam stripping, total investment cost	145
35	Summary of operating costs	146
36	Summary of total investment costs, typical case	147
37	Summary of operating costs, typical case	148
38	Summary of total investment costs, worst case	149
39	Summary of operating costs, worst case	150

LIST OF TABLES

Table Page

1 Results of Spent FCC Catalyst Sample Characterization Tests 30

2 Results of Steam Stripping Experiments B-Series Catalyst, Without Motionless Mixer 34

3 Results of Steam Stripping Experiments C-Series Catalyst, Without Motionless Mixer 36

4 Results of Steam Stripping Experiments D-Series Catalyst, Without Motionless Mixer 39

5 Results of Steam Stripping Experiments E-Series Catalyst, With Motionless Mixer 41

6 Results of Steam Stripping Experiments E-Series Catalyst, Without Motionless Mixer 44

7 Results of Steam Stripping Experiments F-Series Catalyst, With Motionless Mixer 46

8 Results of Steam Stripping Experiments H-Series Catalyst, Without Motionless Mixer 50

9 Results of Steam Stripping Experiments H-Series Catalyst, Without Motionless Mixer 54

10 Results of Steam Stripping Experiments I-Series Catalyst, Without Motionless Mixer 56

11 Results of Experiments Performed to Determine the Effect of Steam Stripping on Coke Volatility, B-Series Catalyst 59

12 Results of Experiments Performed to Determine the Effect of Steam Stripping on Coke Volatility, C-Series Catalyst 61

13 Results of Experiments Performed to Determine the Effect of Steam Stripping on Coke Volatility, E-Series Catalyst 63

14 Results of Experiments Performed to Determine the Effect of Steam Stripping on Coke Volatility, F-Series Catalyst 64

LIST OF TABLES (Continued)

Table		Page
15	Results of Experiments Performed to Determine the Effect of Steam Stripping on Coke Volatility, H-Series Catalyst	65
16	Results of Experiments Performed to Determine the Effect of Steam Stripping on Coke Volatility, H-Series Catalyst	68
17	Results of Coke Volatility Experiments on H-Series Catalyst	71
18	Results of Coke Volatility Experiments on E-Series Catalyst	72
19	Summary of By-Product Formation Experiments, Sulfur and Hydrocarbon Compounds	74
20	Calculated Stripper Off-Gas Analysis C-Series Catalyst	76
21	Calculated Stripper Off-Gas Analysis H-Series Catalyst	77
22	Summary of Results of By-Products Formation Experiments, Ammonia Formation	78
23	Catalyst Activity Tests	79
24	Analysis of Stripper Off-Gas Condensate	86
25	Refinery Wastewater Loadings for Typical Refining Technology	87
26	Steam Stripping Data Regression Analysis	91
27	Steam Stripping Requirement for Sulfur Reduction to 200 vppm	101
28	Recommended Limits of Solids in Boiler Feedwater	125
29	Ionization Constants for the H_2S-Water System at Various Temperatures	129

LIST OF TABLES (Continued)

Table Page

30 Henry's Law Constant for H_2S Versus Temperature 134

31 Capital and Operating Cost Summary for Steam Stripping Concept 144

1. CONCLUSIONS

1. The steam stripping concept identified in Phase I of this program has been tested experimentally in a semi-batch fluidized bed reactor system. Catalysts tested were obtained from existing petroleum refineries in the United States. Altogether, nine catalyst samples were acquired, two appeared to have high water content indicating that they had been partially exposed to air and were no longer representative of an FCC spent catalyst. Consequently, they were eliminated from further experimentation.

2. The spent catalyst samples were exposed to steams at temperatures between 755 and 811 K (900-1000°F), pressures from 1.082×10^5 to 3.427×10^5 Pa (1-35 psig), and steam stripping rates of 1 to 200 kg H_2O/100 kg catalyst. Catalyst steam exposure times ranged from 0 to 3600 seconds.

3. Both air-saturated and oxygen-free distilled waters were used for superheated steam preparation. The presence of oxygen appeared to have no significant effect on sulfur removal efficiency from spent catalyst.

4. The experimental work did not reveal any information that would contradict the conclusions drawn in Phase I of this program. The steam stripping rate of 4 kg H_2O/100 kg of catalyst assumed to evaluate the steam stripping concept economics in Phase I report was further expanded to cover the range between 4 to 100 kg H_2O/100 kg of catalyst. The results are included in this report to allow economics comparisons at those rates.

5. The experimental data demonstrate that reduction of FCC regenerator SO_x emissions is possible via spent catalyst contacting with steam.

   The weight percent sulfur content of coke on spent catalyst samples acquired from petroleum refineries ranged from 0.3 to 1.76%. With no reduction of this sulfur concentration, these catalysts would produce regenerator off-gas containing between 2.43×10^{-4} and 14.49×10^{-4} mole fraction (243-1449 vppm) SO_2.

   Reduction of sulfur on spent FCC catalysts to levels that are equivalent to 2×10^{-4} mole fraction (200 vppm) SO_2 in FCC regenerator off-gas was demonstrated with four catalyst samples. Even though in the case of three samples the equivalent levels of 2×10^{-4} mole fraction (200 vppm) SO_2 were not obtained, a substantial reduction (40-50%) of sulfur on spent catalyst was observed after their exposure to steam.

6. The steam stripping rates needed to reduce sulfur concentrations on spent catalysts to the equivalent levels of 2 x 10^{-4} mole fraction (200 vppm) SO_2 in FCC regenerator off-gas ranged from 2 to 100 kg of steam/100 kg of catalyst.

7. The sulfur removed from spent catalyst appeared in steam in the form of hydrogen sulfide which is readily handled by refineries.

8. Catalyst steam contacting also removed volatilized hydrocarbons from the spent catalyst in addition to sulfur. The gaseous hydrocarbons were identified and included methane, ethane, and propane, with methane prevailing.

   Heavier hydrocarbons were detected as TOC (total organic carbon) condensate in the steam. A linear log-log correlation was found between the steam condensate TOC concentration and steam stripping rate for all catalysts used in this study. This seems to indicate that most of the hydrocarbons are removed from the catalyst in the very initial period of steam contacting and thus the steam stripping may also be used to improve hydrocarbon recoveries in the petroleum refineries. The effect of hydrocarbon removal on catalyst regenerator operation and heat balance will have to be further evaluated for each specific refinery and steam stripping application.

9. In cases of some catalysts, carbonyl sulfide has also been formed in concentrations of about 2×10^{-6} mole fraction (2 vppm) a thousand times lower than those of H_2S. These concentrations are not expected to cause any problems in further processing of H_2S-rich streams.

10. Some formation of ammonia was also observed in steam stripping experiments in concentrations between 3.41×10^{-4} and 5.20×10^{-4} mole fraction (341 to 520 vppm). The formation of ammonia apparently occurs by the same mechanism as that for H_2S. NO_x emissions in the regenerator off-gas may be reduced due to this formation.

11. Exposure to steam at 755-797 K (900-975°F) and 2.39×10^5 Pa (20 psig) for 900 seconds (15 minutes) caused no change in FCC catalyst activity. This is an important observation to assure and maintain minimum interference of steam stripping with present FCC operations in existing refineries.

12. Composition of steam condensate seems to differ very little from compositions of waste waters which refineries presently handle. This suggests that no additional waste water problems, except for increased waste water volume are created as a result of application of steam stripping to sulfur reduction on spent catalyst.

13. A regression analysis of the experimental results from all catalysts tested produced a correlation in which the sulfur removal efficiency is proportional to the product of steam catalyst contact time and steam

stripping rate. The proportionality constant seems to be a function of the catalyst type, form of sulfur on the catalyst, contracting temperature, and design of steam contacting equipment, and has to be experimentally determined. The correlation equation has been modified to describe various designs of steam contacting reactors including semi-fluidized bed, continuous fluidized bed, plug-flow reactor, and counter-current stagewise contacting.

14. The economics of the steam stripping concept were determined as a function of FCC unit size, steam stripping rate, and catalyst attrition rate. The analyses were performed to make their results comparable with the costs for FCC feed desulfurization and add-on processes presented in the Phase I final report. In terms of capital investment costs, (see Figure 34, page 145), steam stripping is competitive with FCC feed desulfurization if the steam stripping rates stay below 70 kg of steam/100 kg of catalyst with an attrition rate of 0.57 kg of catalyst/m^3 of oil (0.2 lb of catalyst/barrel) and below 140 kg steam/100 kg of catalyst in the case of an attrition rate of 0.29 kg/m^3 (0.1 lb/barrel). Operating costs for a catalyst attrition rate of 0.57 kg/m^3 (0.2 lb/barrel) are comparable to those for FCC feedstock desulfurization in the range of 54-81 kg of steam/100 kg of catalyst for 1.84×10^{-2} m^3/s (10,000 barrels per stream day) FCC capacity, 34-49 kg of steam/100 kg of catalyst for 9.20×10^{-2} m^3/s (50,000 barrels per stream day) FCC capacity, and 31-48 kg of steam/100 kg of catalyst for 27.6×10^{-2} m^3/s (150,000 barrels per stream day) FCC capacity. Should the attrition

rate be 0.29 kg/m^3 of oil (0.1 lb/barrel), the operating costs of steam stripping are comparable to those for FCC feedstock desulfurization even if the ranges of steam stripping rates above are doubled (see Figure 35, page 146).

15. Further reduction of the steam stripping costs may result from the use of reduced stripping velocities. The cost estimates presented in this report were calculated based upon 0.61 m/s (2 ft/s) steam superficial linear velocity through the stripper. Reduction of this velocity will proportionally reduce steam consumption. Steam catalyst contact time, however, will have to be increased. This can be easily done by a proper design of the stripper. These cost savings are discussed and demonstrated in Section 6. Additionally, our cost analysis did not include the additional benefits which may result from improved hydrocarbon recovery and recovery of energy from stripping steam condensation. The effects of hydrocarbon removal on FCC regenerator operation, heat balance, and economics have also not been included.

16. Sour operation of the stripping steam condenser may produce streams containing as high as 25% volume H_2S with the condensate not exceeding present water pollution standards for sulfide concentration.

17. The refineries are presently handling sour water produced from several operations (crude distillation, FCC fractionator, coker unit, HDS unit, sulfur plant, etc.). Operations of sour water facilities are similar to the operation of a sour stripping steam condenser. Combining the sour water produced from

steam stripping with sour waters from other operations and treating these waters in one integrated system may further increase the applicability of the steam stripping concept to the refineries.

18. Further substantial improvements in sulfur removal efficiencies may be expected if the steam stripping concept is applied commercially. This statement is based upon observations made in going from a pilot scale to a commercial scale for similar stripping operations. Commercially, the same effects on FCC catalyst were produced by 2 to 5 times lower steam rates than those measured in a pilot scale. Since our experimental results were obtained on a much smaller than pilot scale unit in a semi-batch manner (not identical to commercial operations), even more significant improvements in steam stripping effectiveness to reduce sulfur levels on a spent catalyst should be expected.

2. RECOMMENDATIONS

Based upon the demonstrated ability of the steam stripping concept to reduce sulfur concentrations on spent FCC catalyst with no evident effects on existing FCC unit operation and additional favorable factors which may be realized upon steam stripping operation scale-up, it is highly recommended that this concept be carried through pilot scale development. The pilot scale program should be performed in one of several FCC pilot plants owned by petroleum refinery research centers. This would substantially reduce the cost of such effort.

We also recommend that the pilot scale program should determine the effects of catalyst type, feedstock type, and feedstock pretreatment upon maximum sulfur removal efficiencies via steam stripping. The operating conditions which should be tested upon each of the above combinations include temperature range between 728 and 811 K (850-1000°F), pressure range between 1.082×10^5 and 3.427×10^5 Pa (1-35 psig), and catalyst residence time in stripper between 30 and 900 seconds. Several stripper designs should be tested, including continuous fluidized bed, continuous counter-current stage-wise contactor, and continuous co-current plug-flow contacting. The complete steam analysis for each process stream should be performed to yield complete material and energy balances. The regenerator flue gas should be analyzed to determine the extent of SO_x, NO_x, hydrocarbon, and particulate reduction. The stripper off-gas should be analyzed to determine the concentration of products as a function of operating conditions. Stripper condensate should be characterized to determine the treatability and compatibility of these wastes with other refinery wastes. Finally, based on the results of these tests, new economics should be determined for the steam stripping concept and compared with those for other sulfur reduction techniques.

3. INTRODUCTION

Upon completion of Phase I of EPA Contract No. 68-02-1320, Task 1, Monsanto Research Corporation (MRC) has recommended that several processes be considered as potential candidates for refinery catalytic cracker regenerator SO_x control. The rank ordering of promising processing techniques was established during Phase I and is presented below.

1. Process Modification - steam stripping of spent FCC catalyst

2. Dry Sorption (Westvaco Process and Shell Flue Gas Desulfurization Process)

3. Sodium Sulfite Scrubbing (Wellman-Lord Process)

A detailed discussion and evaluation of each of these techniques was presented in the final report for Phase I. It was determined that steam stripping of FCC catalyst may result in substantial reduction of sulfur compounds deposited on the spent catalyst. This technique would then prevent the sulfur compounds from entering the catalyst regenerator and after their oxidation to sulfur oxides they would be emitted in the regenerator flue gas to the atmosphere.

Based upon the findings in Phase I, Monsanto Research Corporation conducted a laboratory development program and investigated the steam stripping of spent FCC catalyst concept (the processing technique identified in Phase I as the currently most feasible for reducing SO_x emissions from FCC regenerators). Additionally, information was to be acquired to determine economic and environmental aspects of this concept and to establish the needs for further investigation on a pilot scale.

This report summarizes the results and evaluations of experimental work performed during Phase II. After the report conclusions, recommendations, and introduction in Sections 1, 2, and 3, the next section describes the experimental work with experimental apparatus, catalyst samples, and analytical procedures utilized during the program in Section 4.1, and the experimental results and their interpretations in Section 4.2. Process design considerations appear in Section 5. The economic analysis is presented in Section 6.

4. EXPERIMENTAL WORK

During the Phase II experimental program the spent FCC catalyst samples were exposed to various amounts of steam. The experiments were carried out in a semi-batch fluidized catalyst bed reactor. Catalyst samples investigated in the program were obtained from three petroleum refining companies in the United States. The effluent gases from the catalyst testing chamber were analyzed for sulfur and other compounds removed from the catalyst. The description of the experimental equipment, spent catalyst samples, and the analytical techniques and facilities utilized on this program are presented below.

4.1 EXPERIMENTAL EQUIPMENT

4.1.1 Catalyst Test Unit

The test unit, Figure 1, used during this program was designed and fabricated for the purpose of testing catalysts and consisted of the following functional sections:

Reactor

Preparation and metering of simulated process or combustion gases

Effluent gas analysis system

Catalyst handling system

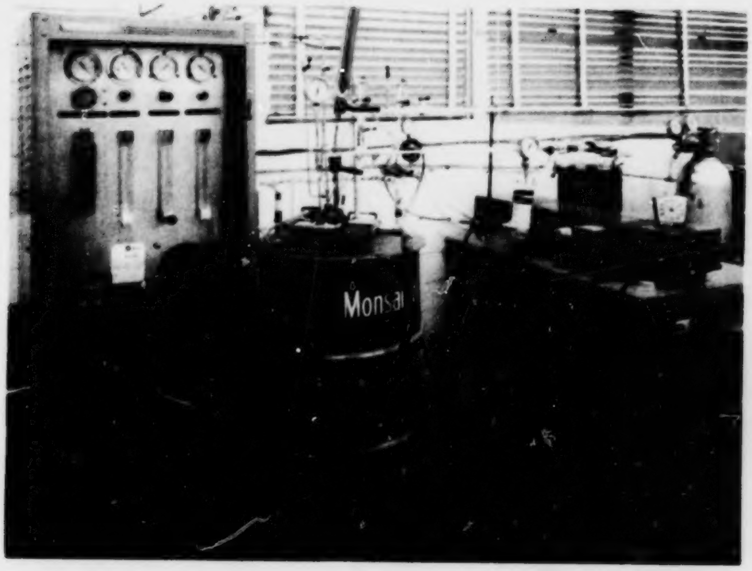

Figure 1. Control panel, reactor, and analysis system

Figure 2 is a schematic flow diagram of the test apparatus. It consisted of a heated reactor chamber mounted in an insulated enclosure. The reactor was designed to operate at temperatures up to 922 K (1200°F) and pressures up to 5.15 x 10^5 Pa (60 psig).

During the experimental work performed on this program, air, nitrogen, and water converted into superheated steam were fed into the reactor charged with FCC catalyst. The gases leaving the reactor were analyzed for compounds stripped and burned off from the catalyst.

For several steam stripping experiments, a motionless mixer was placed inside the reactor chamber (see Figure 3). The mixer was designed to improve gas-catalyst contacting and investigate its effects on catalyst stripping efficiency.

Distilled water was used for superheated steam preparation. For some experiments, the water was deaerated by bubbling prepurified nitrogen through in order to reduce dissolved oxygen content below 2×10^{-5} kg/m^3 (0.02 ppm). The effects of deaerated steam on catalyst stripping efficiency were investigated. The deaeration system with the dissolved oxygen analyzer is shown in Figure 4.

4.1.2 Spent Catalyst Procurement

The primary objective of the catalyst sample procurement was to obtain samples that would represent the catalyst conditions after the catalyst passed through the FCC reactor (spent catalyst) but prior to its regeneration. In addition, samples containing a range of coke concentrations on spent catalyst and sulfur concentrations were needed to establish the effects of these variables on effectiveness of steam stripping.

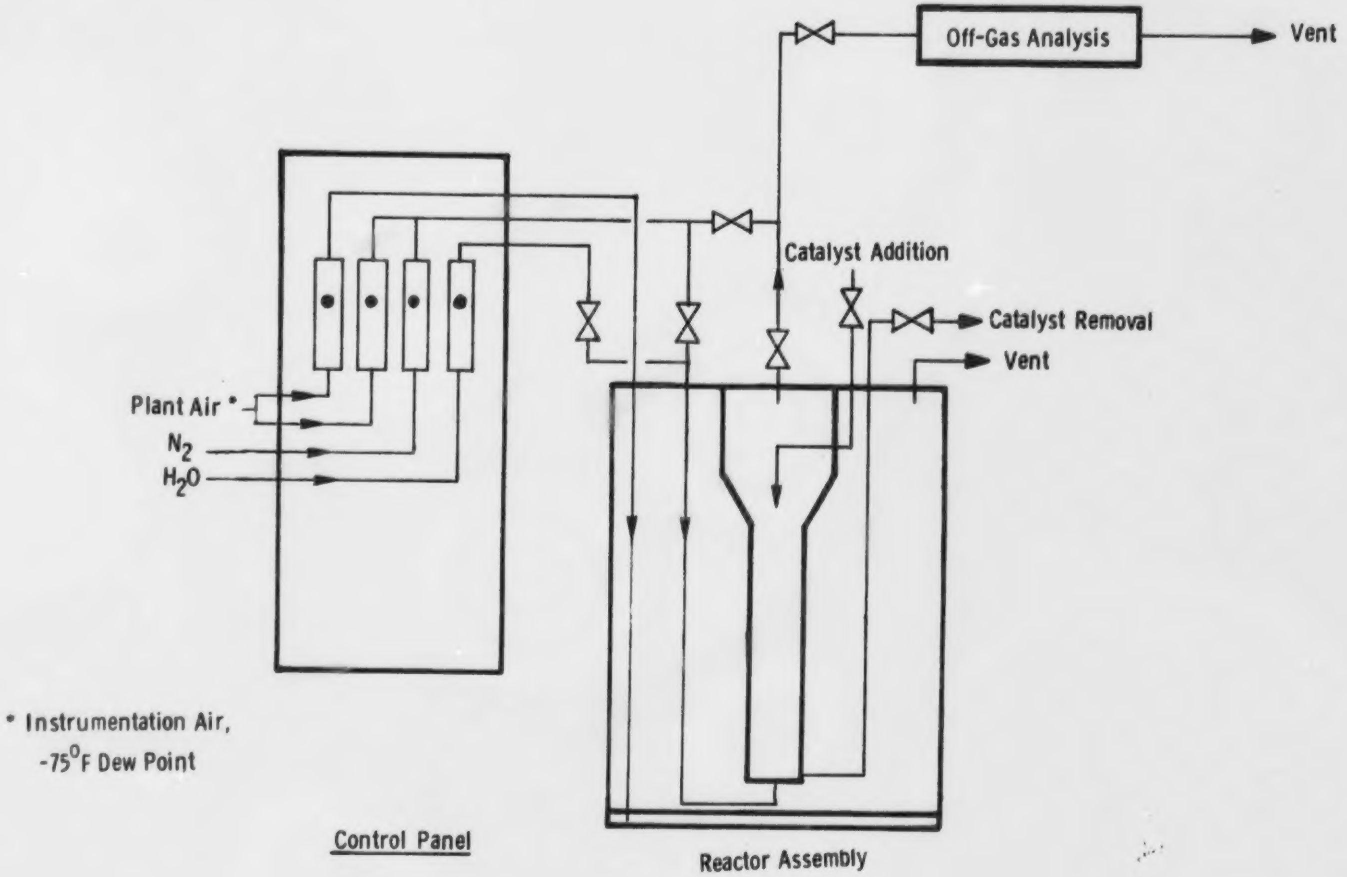
Off-Gas Analysis
Vent
Catalyst Addition
Catalyst Removal
Vent
Plant Air *
N_2
H_2O
* Instrumentation Air,
-75°F Dew Point
Control Panel
Reactor Assembly

Figure 2. Catalyst test unit, schematic flow diagram

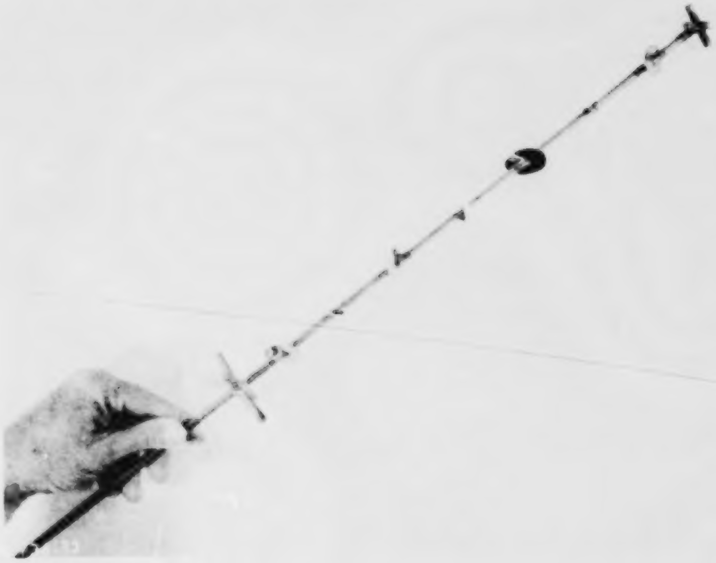

Figure 3. Motionless mixer used in the catalyst reactor

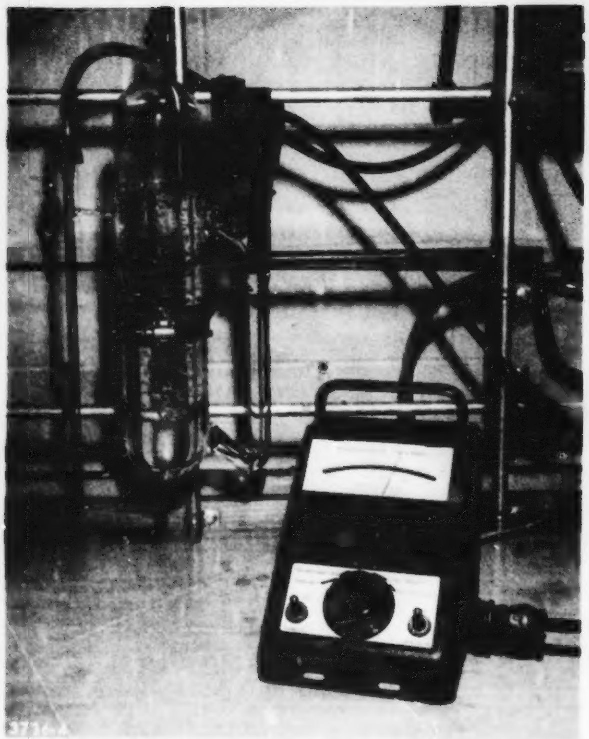

Figure 4. Water deaeration system

The spent FCC catalyst samples used in this program were obtained from various existing petroleum refineries in the U.S. Some refineries asked to supply spent catalyst samples were unable to do so for various reasons; e.g., insufficient manpower for sample collection, lack of appropriate sampling ports, or physical impossibility to collect the sample. Catalyst samples investigated in this study were supplied by three American oil companies.

The catalyst samples were delivered in the following amounts.

Catalyst	m^3	Gallons
"A"	1.89×10^{-2}	5
"B"	0.76×10^{-2}	2
"C"	1.89×10^{-2}	5
"D"	0.38×10^{-2}	1
"E"	1.89×10^{-2}	5
"F"	1.89×10^{-2}	5
"G"	20.80×10^{-2}	55
"H"	20.80×10^{-2}	55
"I"	0.57×10^{-2}	1.5

The "A" sample was shipped in an open top can inside a disposable plastic garbage bag. The "G" sample was shipped in an open top drum. Upon arrival, the lid on both containers was not securely fastened.

The samples were characterized according to the procedures described in Section 4.1.4. The results of these analyses are presented in Table 1 (Section 4.2) and indicate that the samples contained about 2.5% and 5%+ moisture, respectively.

Evidently the samples absorbed moisture from the air, lost their integrity and were not representative of spent FCC catalyst. Consequently, they were eliminated from further experimentation. Samples "B", "C", "D", "E", "F", "H", and "I" were shipped in airtight containers and were used in our experiment.

4.1.3 Catalyst Handling

As discussed in the Phase I final report, the cracking catalysts in use today are primarily of the zeolitic type and have an affinity for water. In the FCC unit operations the catalyst is exposed to elevated temperatures, 811-922 K (1000-1200°F). At these temperatures the catalyst is essentially dry. Any exposure of the spent catalyst to water vapor or oxidation atmosphere of ambient air could result in water absorption and also yield slow oxidation of hydrocarbons deposited on the spent catalyst. Both of these phenomena would make the sample non-representative of spent FCC catalyst. Furthermore, the stripping reactions could occur under these conditions at a slow rate, removing some sulfur and hydrocarbons before the steam stripping experiments. Presence of air could also cause some side reactions with the hydrocarbons and change the nature of the coke originally present on the spent catalyst. In order to maintain the spent catalyst sample integrity it was imperative to prevent catalyst exposure to water vapor or air during the sample collection, shipment, and handling in the laboratory.

We advised the petroleum refineries who supplied the spent catalyst that the samples should be collected in an inert atmosphere (such as nitrogen) and shipped in an airtight container under a nitrogen blanket.

Determination of whether or not the samples were shipped to MRC in airtight containers was made at the time the samples arrived at our location. Each shipping container was inspected for leaks and the catalyst from any container not securely sealed was not used during our research program. In all of our efforts, we tried to use only those catalyst samples which were representative of the spent catalyst in an FCC unit. The characterization analyses were an additional measure of the spent catalyst sample representativeness. It should be noted, however, that we do not know whether or not each sample maintained integrity because of the thermal cycling which occurred during the sample collection, shipment, and experimentation. Each sample had to be collected at an elevated temperature of 755 to 811 K (900 to 1000°F), cooled down for shipment, and reheated during our experimentation program.

In order to maintain catalyst integrity during our experimentation a system was devised to handle the spent catalyst samples without contamination. A flow capability of the spent FCC catalyst was utilized in this system. The spent catalyst sample container (a can or drum) was pressurized with nitrogen to about 4.754×10^7 Pa (0.5 psig). This pressure was maintained at all times to prevent air or moisture leakage into the container. A vent pipe made of 9.53×10^{-3} m (3/8 inch) stainless steel tubing was inserted into the container for transferring the catalyst into a preweighed nitrogen purged flask (see Figure 5). A stainless steel ball valve with Teflon seals was used to control catalyst flow out of the catalyst sample container. The weighing flask was fitted with pinch clamps to prevent air and moisture leakage to the catalyst contained inside the flask. Typically, the flask was charged with 0.4 kg of spent catalyst.

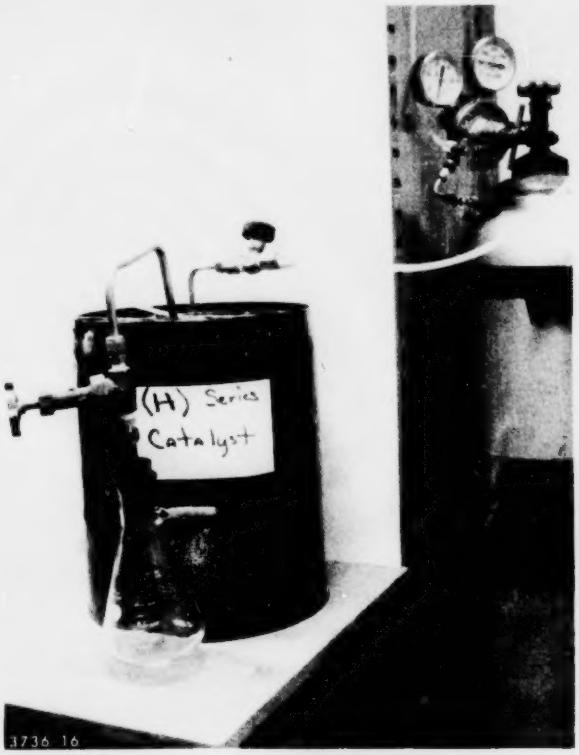
(H) Series
Catalyst
3736 16

Figure 5. Photograph of catalyst storage container with charging flash attached

The catalyst was then transferred to a nitrogen purged charging bomb (see Figure 6). This bomb was used to transfer a spent FCC catalyst sample to a preheated reactor. Actual charging was done under a nitrogen blanket and 4.81×10^5 Pa (55 psig) pressure according to the diagram shown in Figure 7.

When the catalyst charging was completed, the charging bomb was remounted onto the weighing flask to remove all the residual catalyst retained in the bomb. The difference in weight of the flask before and after the catalyst charging operation was a measure of the amount of the catalyst sample charged into the reactor.

The fluid nature of the spent FCC catalyst was utilized for the quantitative removal of catalyst from the catalyst unit. The catalyst was pneumatically transported to a 10^{-3} m^3 (1000 ml) Erlenmeyer collection flask using nitrogen as the carrier gas. The catalyst was discharged through the catalyst removal tubing (see Figure 2).

4.1.4 Steam Stripping Experiments and Catalyst Characterization

The procedures used to perform the spent FCC catalyst steam stripping experiments and catalyst characterization tests are presented below.

4.1.4.1 Catalyst Steam Stripping -

A known weight of catalyst sample (0.35-0.40 kg) was placed into the catalyst reactor. The catalyst was then heated to the predetermined steam stripping temperature. After the reactor reached the desired temperature, a known quantity of superheated steam was passed through the catalyst bed at a

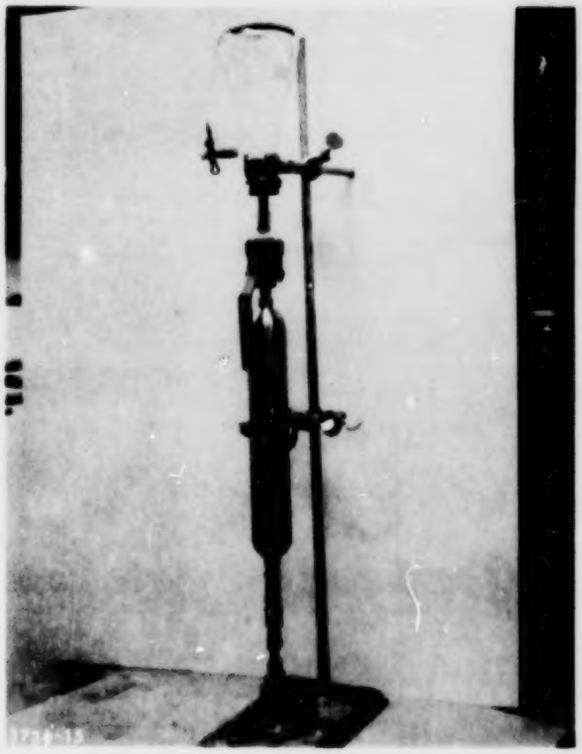

Figure 6. Weighing flask mounted on catalyst charging bomb

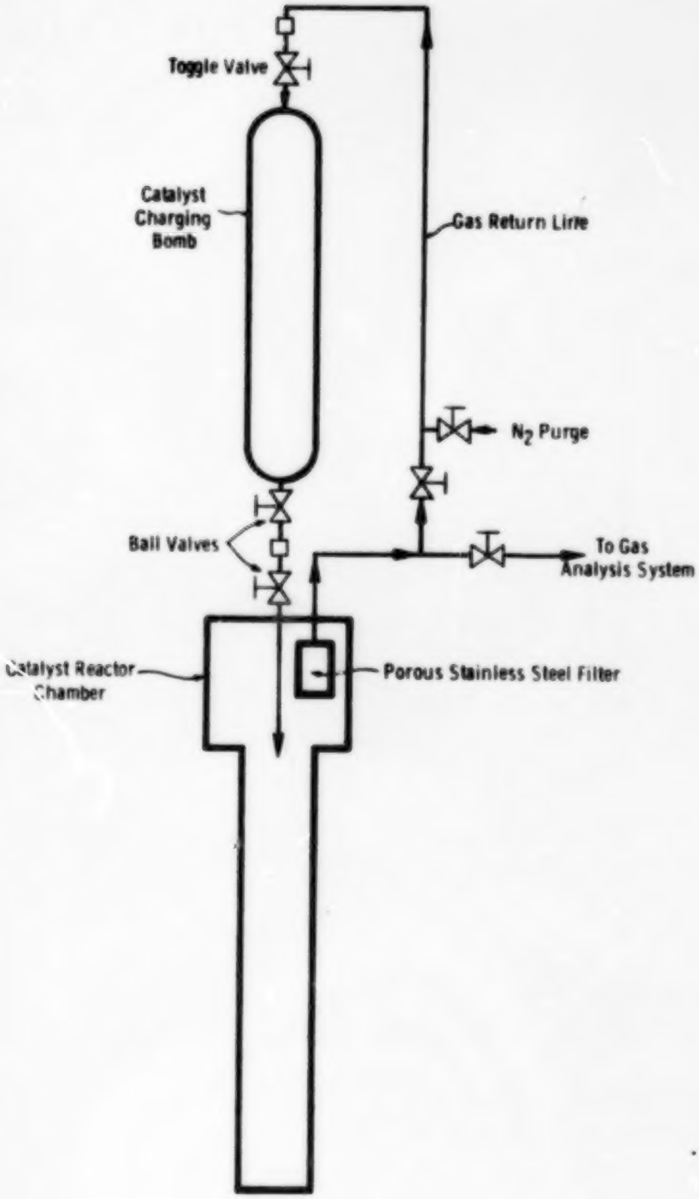
Toggle Valve
Catalyst Charging Bomb
Gas Return Line
N2 Purge
Ball Valves
To Gas Analysis System
Catalyst Reactor Chamber
Porous Stainless Steel Filter

Figure 7. Charging of the catalyst into a hot reactor

controlled rate which maintaned the catalyst bed in a fluidized state. The entire off-gas stream was sent through the H_2S analysis system where the sulfur content of the stripping steam was determined. After the steam stripping experiments, the H_2S analysis system was replaced with $SO_2/SO_3/H_2SO_4$ analysis system and the same catalyst charge exposed to air to determine tha residual sulfur remaining on the catalyst according to the procedure described in the following section.

4.1.4.2 Determination of the Sulfur Content of Coke -

A known weight of catalyst was charged into the catalyst reactor (0.35-0.40 kg) and oxidized with air at 922 K (1200°F). The temperature of the catalyst bed was controlled by adjusting the flow rate of air through the reactor. The entire gas stream leaving the reactor was analyzed for oxidation products of sulfur, $SO_2/SO_3/H_2SO_4$. From the results of this analysis the total sulfur of spent catalyst coke was calculated.

4.1.4.3 Determination of the H_2O Content of Spent FCC Catalyst -

A known weight of each spent catalyst type (0.35-0.40 kg) was charged to the catalyst reactor in an inert nitrogen atmosphere as described in Section 4.1.3. A slow, 1.67×10^{-5} m^3/s (1 liter/minute) nitrogen purge through the catalyst at bed was used to remove water vapor desorbed from the catalyst 783 K (950°F). This test was performed for a standard period of 300 s (5 minutes) to prevent an excessive volatilization of other compounds from the catalyst. The off-gases were collected in a preweighed drying tube filled with silica gel. The weight difference of this tube before and after the test was a measure of a catalyst moisture content.

4.1.4.4 Determination of the Coke Content of Spent FCC Catalyst -

A known weight of each spent catalyst type (0.35 - 0.40 kg) was charged in the catalyst reactor. The coke deposits were oxidized with air at 922 K (1200°F). Upon completion of the coke combustion the catalyst was removed from the test unit. The difference in weight of catalyst before and after the test was the measure of the coke and the moisture content of the catalyst. Subtracting the moisture content determined according to the procedure described in the previous paragraph produced the value for the catalyst coke content. This value would, of course, be representative of all compounds forming the coke (carbon, hydrogen, sulfur, nitrogen, oxygen, etc.).

4.1.5 Other Experiments

Analytical procedures were developed to determine the various by-products formed during steam stripping of FCC catalyst. Specifically, we analyzed the off-gases from the steam stripping experiments for carbonyl sulfide, carbon disulfide, mercaptan sulfur, ammonia, and hydrocarbons including methane, ethylene, propane, propylene, and benzene. The analytical procedures used to analyze all these compounds will be described in the following section. The purpose of these tests was to determine the maximum concentration of products formed by pyrolyzing the coke on spent catalyst. The tests identified the types of compounds to be analyzed for in subsequent experiments and determined the maximum possible decrease in coke content caused by heating.

4.1.6 Analysis

Wherever possible, standard and well established analytical techniques were utilized. This was the case with $SO_2/SO_3/H_2SO_4$, H_2S, and NH_3 analyses. EPA Method #8 sampling train and procedure for "Determination of Sulfuric Acid Mist and Sulfur Dioxide Emissions from Stationary Sources" was used for determination of oxidized sulfur compounds (Federal Register, Vol. 36, No. 247, December 23, 1971, pp. 24893-24895). An EPA Method #11 sampling train for "Determination of Hydrogen Sulfide Emissions from Stationary Sources" (Federal Register, Vol. 39, No. 47, March 8, 1974, pp. 9321-9323) was used for hydrogen sulfide determination. This method was modified in order to quantitatively condense and collect superheated steam. The normal procedure requires that midget impingers with 0.05×10^{-3} m^3 (50 ml) capacity be utilized. For this program, standard 0.5×10^{-3} m^3 (500 ml) Greenburg-Smith impingers were used.

The ammonia analysis of the stripping steam condensate was made according to the procedures outlined in Standard Methods for the Examination of Water and Wastewater (New York, American Public Health Association, 13th Edition, 1971, Procedure #132).

4.1.6.1 Sulfur Compound and Hydrocarbon Determination -

The analysis for sulfur containing compounds and hydrocarbons was performed on steam condensate samples as well as gas samples collected into the Tedlar bag during the steam stripping experiments. All samples were analyzed in the F&M Model 720 chromatograph utilizing a dual column and detection system. A Porapak-Q gas chromatographic column was used to separate the various components of the sample analyzed. The chromatographic system employed a temperature programming feature to aid in separation of hydrocarbons and sulfur compounds. The detector system considted of a Tracor 1 sulfur selective flame photometric detector and a hydrocarbon flame ionization detector. Both detectors were used simultaneously to detect hydrocarbons and sulfur compounds.

4.1.6.2 Analysis for the Total Organic Carbon Content of Stripper Condensate -

In order to determine the extent of hydrocarbon contamination of the stripping steam condensate the entire stripper off-gas stream was condensed and quantitatively analyzed for carbon. The samples were analyzed according to Procedure #138 in the Standard Methods for the Examination of Water and Wastewater using the Beckman TOC analyzer.

4.1.6.3 Volatility of Coke on Spent Catalyst -

In this test, the catalyst samples were placed in an evacuated quartz tube sealed on one end and heated from room temperature to 811 K (1000°F). The other end of the tube was connected to the mass spectrometer (Du Pont CEC Model 21-103 C). Gases evolved during the test were introduced into the instrument for the analysis.

4.2 EXPERIMENTAL RESULTS

4.2.1 Catalyst Characterization

Each catalyst sample was characterized immediately before it was submitted to steam stripping experiments in order to minimize potential changes in catalyst samples due to aging over long periods of time. Each test was performed at least three times and an average value was then calculated, except for sample "D" which had only one characterization analysis performed because of the sample limited amount. The results for all catalysts obtained from the petroleum refineries are presented in Table 1. The values in Table 1 are presented as averages with plus/minus percent deviations observed when the multiple characterization tests were performed. The sulfur content of coke deposited on "as-received" catalyst was calculated based on an average value of catalyst coke content.

The data in Table 1 were used to calculate the equivalent regenerator SO_2 emissions from the FCC regenerator. A math model was developed during Phase I of this program (see Appendix G in the Phase I final report) to predict the FCC regenerator SO_2 emissions after calculating the carbon, hydrogen, and sulfur contents of the coke and the CO_2/CO ratio of the gases leaving the regenerator. This model was applied to convert the sulfur content of coke to the equivalent regenerator SO_2 concentrations. The following assumptions were made in using this conversion method:

Hydrogen content of coke, $H = 0.1$ (weight fraction)
CO_2/CO ratio, $R = 1$ (mole ratio)

According to the math model, the weight fraction of sulfur in coke before combustion would have to be 0.00243 in order to obtain a concentration of 2×10^{-4} mole fraction (200 vppm) SO_2 in the regenerator off-gas with $R = 1$ and $H = 0.1$. (This can also be determined from the diagram in Figure 5, page 27 in the Phase I final report, which was produced based on the equation in Appendix G). The calculated

Table 1. RESULTS OF SPENT FCC CATALYST SAMPLE CHARACTERIZATION TESTS

Sample	Moisture (%)	Coke (%)	Sulfur (%)	SO_2 concentration (mole fraction)
A	2.488±0.763	0.489±4.3	0.451±2.0	3.71×10^{-6}
B	1.260±3	0.663±4.7	1.76±3.0	14.49×10^{-6}
C	0.617±8.4	1.202±3.2	1.013±0.6	8.34×10^{-6}
D	0.363	1.889	0.624	5.14×10^{-6}
E	0.0	1.016±3.3	0.520±2.5	4.28×10^{-6}
F	0.0628±13	0.854±1.55	0.595±17	4.90×10^{-6}
G	5+	N.A.	N.A.	---
H	0.492±4.4	1.304±2.1	0.295±2.3	2.43×10^{-6}
I	0.680±24	1.529±5.52	0.748±6.45	6.16×10^{-6}

initial equivalent regenerator SO_2 concentrations are also included in Table 1 for each of the catalyst samples used in this program. An identical procedure was used to predict the equivalent regenerator SO_2 concentrations after the steam stripping experiments.

As shown in Table 1, samples "A", "B", and "G" contained 2.49%, 1.26%, and 5+% (by weight) of moisture, respectively. It was felt that samples "A" and "G" had lost their integrity due to use of improper containers for their collection and shipment (refer to Section 4.1.2). They were thus eliminated from any further experimentation. Although sample "B" contained 1.26% (by weight) of moisture, it was believed that this moisture content was marginal and that this sample should undergo experimentation.

4.2.2 Catalyst Steam Stripping

The steam stripping experiments were performed according to the procedure described in Section 4.1.4. Various spent catalyst samples received from petroleum refineries have been exposed to steam at different temperatures, pressures, steam stripping rates, and catalyst residence times in the stripper reactor. The residence times were expressed by two variables, steam-catalyst contact time and catalyst-steam exposure time.

The steam-catalyst contact time is defined as the length of time that the steam is in contact with the catalyst. It is calculated by dividing the fluidized catalyst height by the steam superficial linear velocity and correcting for bed porosity.

Catalyst-steam exposure time is defined as the length of time which the catalyst is located in a steam environment. More accurately, it is the catalyst residence time in the stripper.

The following are the ranges of the variables studied during this program:

Stripping temperature,	K	755-811
	°F	900-1000
Stripping pressure,	Pa	$1.08x10^{5}-3.43x10^{5}$
	psig	1-35
Steam stripping rate,	kg H_2O/100 kg catalyst	1-200
Steam-catalyst contact time,	s	0.5-2.0
Catalyst-steam exposure time,	s	0-3600

After a catalyst sample steam stripping experiment and after sulfur analysis of the steam stream, the catalyst sample was oxidized in air and the effluent stream analyzed for sulfur oxidation compounds. Thus, knowing the original concentration of sulfur on the catalyst charged to the reactor, a good sulfur balance check could be made by adding total sulfur stripped from the catalyst with the steam and total sulfur remaining on the catalyst and removed after the oxidation with air.

All the experimental and analytical data were recorded on specially prepared blank forms. An example of this form is presented in Appendix A. It includes calculation procedures used to determine final results.

The final results of all steam stripping experiments have been summarized in tabular form and are presented in Table 2 through 10. Each table identifies clearly the conditions at which the experiments were carried out: temperature, pressure, catalyst source, and other measured variables including steam stripping rate, steam-catalyst contact time, superficial steam catalyst contact time, catalyst-steam exposure time, stripper superficial velocity, and oxygen content of superheater feed water. The tables also list the calculated results of percent sulfur removal by steam stripping and equivalent regenerator SO_2 concentrations. For convenience and to better observe the effect of steam stripping rate on sulfur removal, the data are also presented in graphic form (Figures 8 through 16). Here, steam stripping rate has been plotted against the equivalent regenerator SO_2 concentration. The experiments identified as not shown on graphs were obtained at steam stripping rates that are higher than those presented on graphs. For results of these experiments refer to corresponding experimental data tables.

4.2.3 Hydrocarbon Volatilization During Steam Stripping

A number of tests were performed to determine the amount of hydrocarbons that would volatilize during steam stripping experiments and contaminate the steam condensate. The procedure followed in these tests was described in Section 4.1.6.3. The results are summarized in Tables 11 through 16. For convenience, the experiment operating conditions are also included in the data tables.

Table 2. RESULTS OF STEAM STRIPPING EXPERIMENTS

B-Series Catalyst, Without Motionless Mixer

	Experiment						
	B-3	B-5	B-7	B-8	B-9	B-11	B-13
Catalyst bed temperature, (K) (°F)	811 1000	811 1000	811 1000	811 1000	811 1000	811 1000	811 1000
Steam stripping rate (SSR), (kg H_2O/100 kg catalyst)	5.31	32.8	17.6	39.8	9.01	21.9	1372
Stripper pressure, (Pa) (psig)	$1.08x10^5$ 1.0	$1.08x10^5$ 1.0	$2.39x10^5$ 20	$2.39x10^5$ 20	$3.08x10^5$ 30	$3.43x10^5$ 35	$3.43x10^5$ 35
Steam-catalyst contact time, (s)	0.123	0.837	0.145	0.156	0.212	0.192	0.0164
Superficial steam-catalyst contact time, T_S, (s)	1.23	0.791	1.36	1.48	2.05	1.81	0.155
Catalyst-steam exposure time, (s)	180	720	300	738	180	360	1920
Stripper superficial velocity, (m/s) (ft/s)	 0.276 0.907	 0.408 1.34	 0.155 0.509	 0.197 0.646	 0.134 0.438	 0.158 0.517	 0.140 0.458
O_2 content of superheater feedwater, (kg/m^3)	$2.0x10^{-5}$	$2.0x10^{-5}$	$2.0x10^{-5}$	$2.0x10^{-5}$	$2.0x10^{-5}$	$2.0x10^{-5}$	$2.0x10^{-5}$
Sulfur removal (% by wt)	32.4	49.5	34.1	50.2	44.4	46.9	76.3
Residual equivalent regenerator SO_2 concentration, S_O, (mole fraction) (vppm)	 $9.79x10^{-4}$ 979	 $7.32x10^{-4}$ 732	 $9.51x10^{-4}$ 951	 $7.21x10^{-4}$ 721	 $8.05x10^{-4}$ 805	 $7.69x10^{-4}$ 769	 $3.44x10^{-4}$ 344

Remarks: Equivalent regenerator SO_2 concentration before stripping, $14.49x10^{-4}$ mole fraction SO_2
1449 vppm SO_2(S_1)

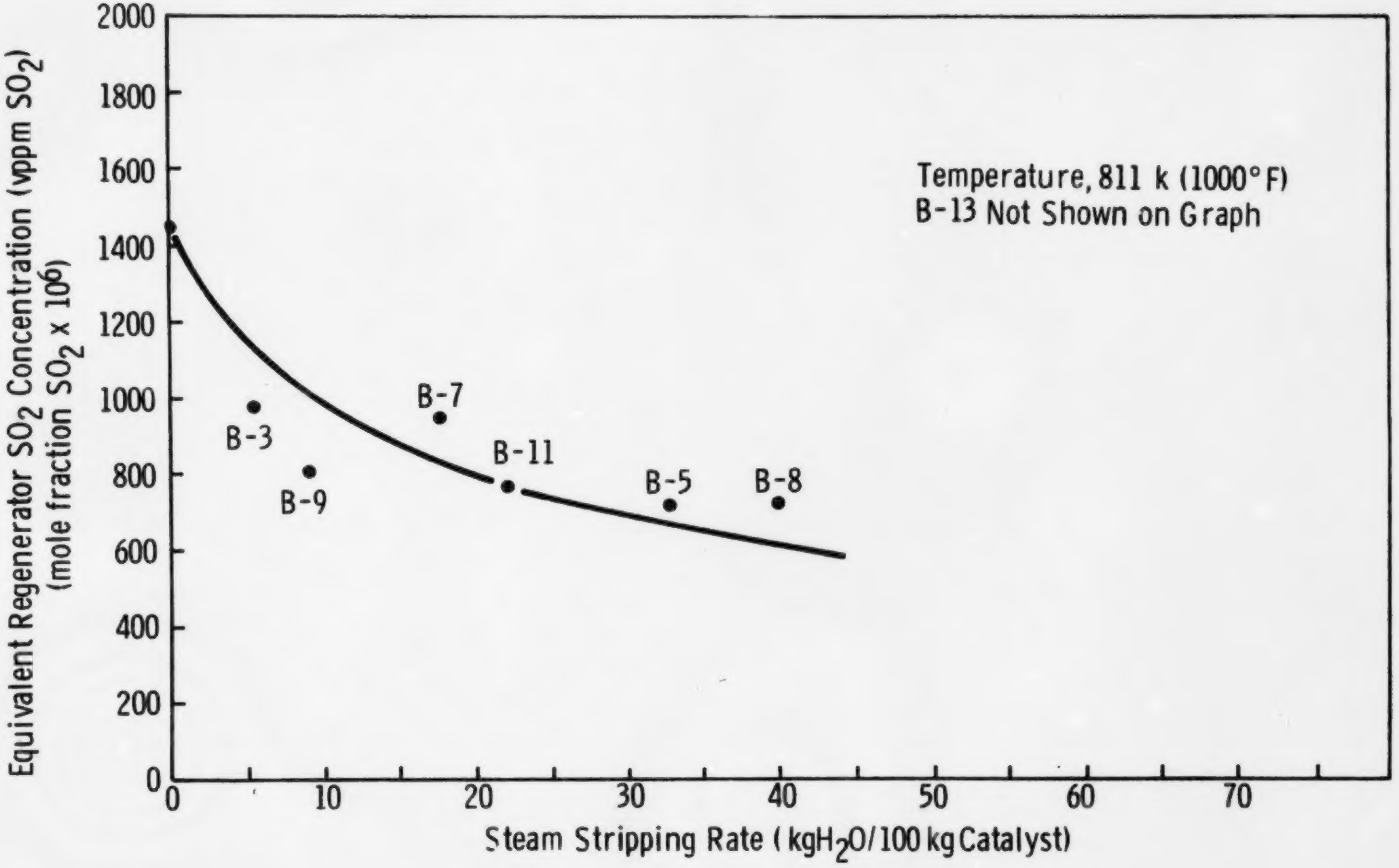

Figure 8. Results of steam stripping experiments on B-Series catalyst

Table 3. RESULTS OF STEAM STRIPPING EXPERIMENTS

C-Series Catalyst, without Motionless Mixer

	Experiment						
	C-6	C-7	C-8	C-9	C-10	C-11	C-12
Catalyst bed temperature, (K)	811	811	811	811	811	811	811
(°F)	1000	1000	1000	1000	1000	1000	1000
Steam stripping rate (SSR), (kg H_2O/100 kg catalyst)	7.31	14.1	32.6	8.08	16.0	5.98	37.2
Stripper pressure, (Pa)	1.08×10^5	1.08×10^5	1.08×10^5	2.39×10^5	2.39×10^5	2.39×10^5	2.39×10^5
(psig)	1.0	1.0	1.0	20	20	20	20
Steam-catalyst contact time, (s)	0.0939	0.0971	0.0842	0.187	0.191	0.134	0.164
Superficial steam-catalyst contact time, T_S, (s)	0.888	0.924	0.798	1.78	1.80	1.27	1.54
Catalyst-steam exposure time, (s)	180	360	720	180	360	95	720
Stripper superficial velocity, (m/s)	0.396	0.408	0.469	0.211	0.212	0.258	0.199
(ft/s)	1.30	1.34	1.54	0.692	0.695	0.845	0.652
O_2 content of superheater feedwater, (kg/m³)	2.0×10^{-5}	2.0×10^{-5}	2.0×10^{-5}	2.0×10^{-5}	2.0×10^{-5}	2.0×10^{-5}	2.0×10^{-5}
Sulfur removal (% by wt)	29.4	34.2	35.2	36.0	40.9	26.8	35.1
Residual equivalent regenerator SO_2 Concentration, S_O, (mole fraction)	5.92×10^{-4}	5.49×10^{-4}	5.40×10^{-4}	5.34×10^{-4}	4.93×10^{-4}	6.10×10^{-4}	5.41×10^{-4}
(vppm)	592	549	540	534	493	610	541

Remarks: Equivalent regenerator SO_2 concentration before stripping, 8.34×10^{-4} mole fraction SO_2
834 vppm SO_2 (S_1)

Table 3 (continued). RESULTS OF STEAM STRIPPING EXPERIMENTS

C-Series Catalyst, without Motionless Mixer

	C-14	C-15	C-17	C-18	C-19	C-20	C-21
Catalyst bed temperature, (K)	811	811	811	811	811	811	811
(°F)	1000	1000	1000	1000	1000	1000	1000
Steam stripping rate (SSR), ($kg\ H_2O$/100 kg catalyst)	29.5	7.59	219	109	467	738	631
Stripper pressure, (Pa)	$3.43x10^5$	$3.43x10^5$	$2.39x10^5$	$3.43x10^5$	$1.08x10^5$	$2.39x10^5$	$3.43x10^5$
(psig)	35	35	20	35	1.0	20	35
Steam-catalyst contact time, (s)	0.286	0.185	0.176	0.258	0.0295	0.0404	0.0657
Superficial steam-catalyst contact time, T_S, (s)	2.69	1.75	1.48	2.43	0.279	0.389	0.631
Catalyst-steam exposure time, (s)	720	120	4050	2400	3600	3600	3600
Stripper superficial velocity, (m/s)	0.135	0.215	0.220	0.154	0.421	0.205	0.143
(ft/s)	0.444	0.704	0.722	0.505	1.38	0.673	0.469
O_2 content of superheater feedwater, (kg/m)	$2.0x10^{-5}$	$2.0x10^{-5}$	$2.0x10^{-5}$	$2.0x10^{-5}$	$2.0x10^{-5}$	$2.0x10^{-5}$	$2.0x10^{-5}$
Sulfur removal (% by wt)	36.0	33.0	44.9	42.1	60.7	56.0	43.5
Residual equivalent regenerator SO_2 concentration, S_O, (mole fraction)	$5.34x10^{-4}$	$5.59x10^{-4}$	$4.60x10^{-4}$	$4.83x10^{-4}$	$3.27x10^{-4}$	$3.67x10^{-4}$	$4.71x10^{-4}$
(vppm)	534	559	460	483	327	367	471

Remarks: Equivalent regenerator SO_2 concentration before stripping, $8.34x10\ ^{4}$ mole fraction SO_2
834 vppm SO_2 (S_1)

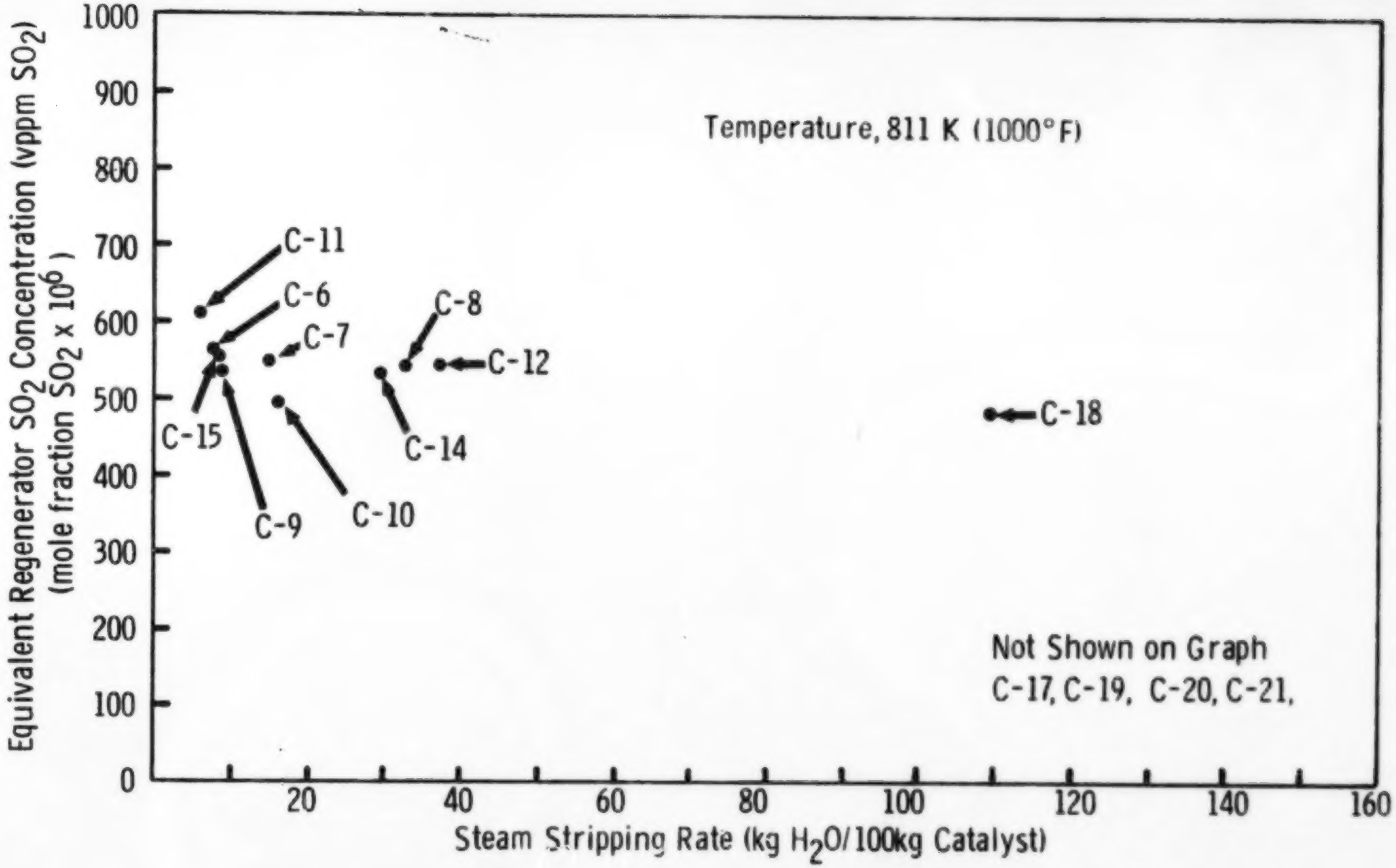

Figure 9. Results of steam stripping experiments on C-Series catalyst

Table 4. RESULTS OF STEAM STRIPPING EXPERIMENTS

D-Series Catalyst, without Motionless Mixer

	Experiment			
	D-2	D-4	D-5	D-7
Catalyst bed temperature, (K)	811	811	811	811
(°F)	1000	1000	1000	1000
Steam stripping rate (SSR), (kg H_2O/100 kg catalyst)	17.7	41.1	90.5	16.6
Stripper pressure, (Pa)	2.39×10^5	2.39×10^5	2.39×10^5	2.39×10^5
(psig)	20	20	20	20
Steam-catalyst contact time, (s)	0.173	0.149	0.140	0.184
Superficial steam-catalyst contact time, T_s, (s)	1.63	1.40	1.32	1.73
Catalyst-steam exposure time, (s)	360	720	.500	3600
Stripper superficial velocity, (m/s)	0.201	0.201	0.201	0.201
(ft/s)	0.659	0.661	0.661	0.661
O_2 content of superheater feedwater, (kg/m)	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$
Sulfur removal (% by wt)	25.0	64.3	57.8	67.3
Residual equivalent regenerator SO_2 concentration, So (mole fraction)	3.85×10^{-4}	1.83×10^{-4}	2.17×10^{-4}	1.68×10^{-4}
(vppm)	385	183	217	168

Remarks: Equivalent regenerator SO_2 concentration before stripping, 5.13×10^{-4} mole fraction SO_2
513 vppm $SO_2(S_1)$

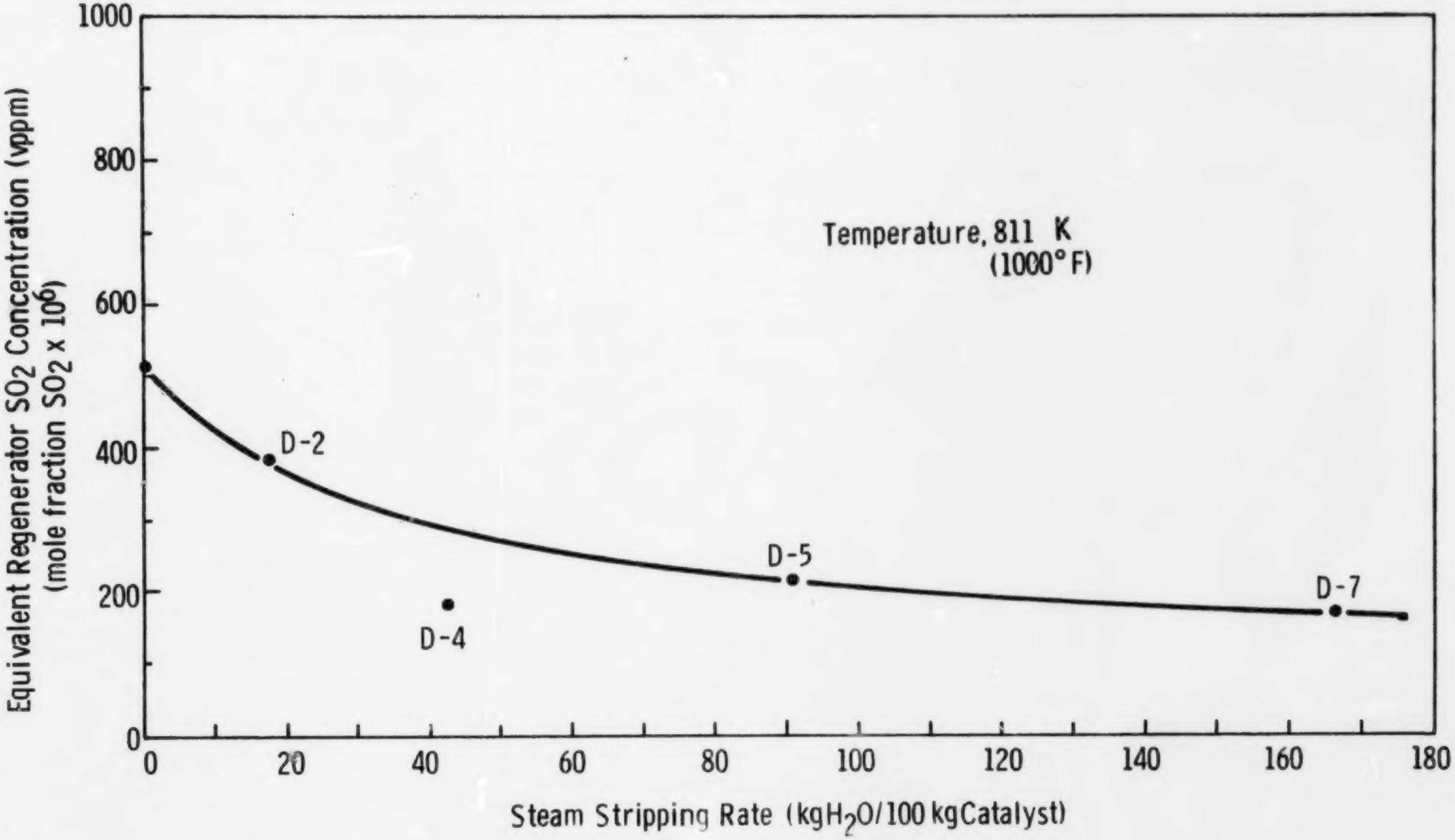

Figure 10. Results of steam stripping experiments on D-Series catalyst

Table 5. RESULTS OF STEAM STRIPPING EXPERIMENTS

E-Series Catalyst, with Motionless Mixer

	Experiment				
	E-6	E-8	E-15	E-16	E-17
Catalyst bed temperature, (K)	853	811	811	811	811
(°F)	1075	1000	1000	1000	1000
Steam stripping rate (SSR), (kg H_2O/100 kg catalyst)	*97.8	148	*6.32	13.3	5.2
Stripper pressure, (Pa)	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5
(psig)	1.0	1.0	1.0	1.0	1.0
Steam-catalyst contact time, (s)	0.095	0.0639	0.180	0.162	0.189
Superficial steam-catalyst contact time, T_s, (s)	0.88	0.584	1.630	1.48	1.84
Catalyst-steam exposure time, (s)	2400	2400	260	546	264
Stripper superficial velocity, (m/s)	0.411	0.442	0.218	0.256	0.197
(ft/s)	1.35	1.45	0.716	0.840	0.645
O_2 content of superheater feedwater, (kg/m³)	$<9.1 \times 10^{-5}$	$<9.1 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$
Sulfur removal (% by wt)	62.5	38.8	15.4	28.7	15.4
Residual equivalent regenerator SO_2 concentration, S_o, (mole fraction)	1.62×10^{-4}	2.27×10^{-4}	3.62×10^{-4}	3.05×10^{-4}	3.61×10^{-4}
(vppm)	162	227	362	305	361

Remarks: Equivalent regenerator SO_2 concentration before stripping, 4.28×10^{-4} mole fraction on SO_2
428 vppm SO_2 (S_1)

* Estimated values

Table 5 continued. RESULTS OF STEAM STRIPPING EXPERIMENTS

E-Series Catalyst, with Motionless Mixer

	Experiment				
	E-18	E-21	E-22	E-23	E-24
Catalyst bed temperature, (K) (°F)	811 1000	811 1000	811 1000	811 1000	811 1000
Steam stripping rate (SSR), (kg H_2O/100 kg catalyst)	2.85	11.6	1.34	6.16	11.8
Stripper pressure, (Pa) (psig)	1.08×10^5 1.0	1.08×10^5 1.0	1.08×10^5 1.0	1.08×10^5 1.0	1.08×10^5 1.0
Steam-catalyst contact time, (s)	0.113	0.0989	0.0708	0.0942	0.0871
Superficial steam-catalyst contact time, T_s, (s)	1.06	0.937	0.642	0.879	0.826
Catalyst-steam exposure time, (s)	84	300	24	150	270
Stripper superficial velocity, (m/s) (ft/s)	0.347	0.399	0.555	0.418	0.445
O_2 content of superheater feedwater, (kg/m^3)	2.0×10^{-5}	2.0×10^{-5}	2.0×10^{-5}	2.0×10^{-5}	2.0×10^{-5}
Sulfur removal (% by wt)	20.8	20.8	8.17	14.0	14.6
Residual equivalent regenerator SO_2 concentration, S_0, (mole fraction) (vppm)	3.62×10^{-4} 362	3.39×10^{-4} 339	3.93×10^{-4} 393	3.68×10^{-4} 368	3.66×10^{-4} 366

Remarks: Equivalent regenerator SO_2 concentration before stripping, 4.28×10^{-4} mole fraction on SO_2
428 vppm SO_2 (S_1)

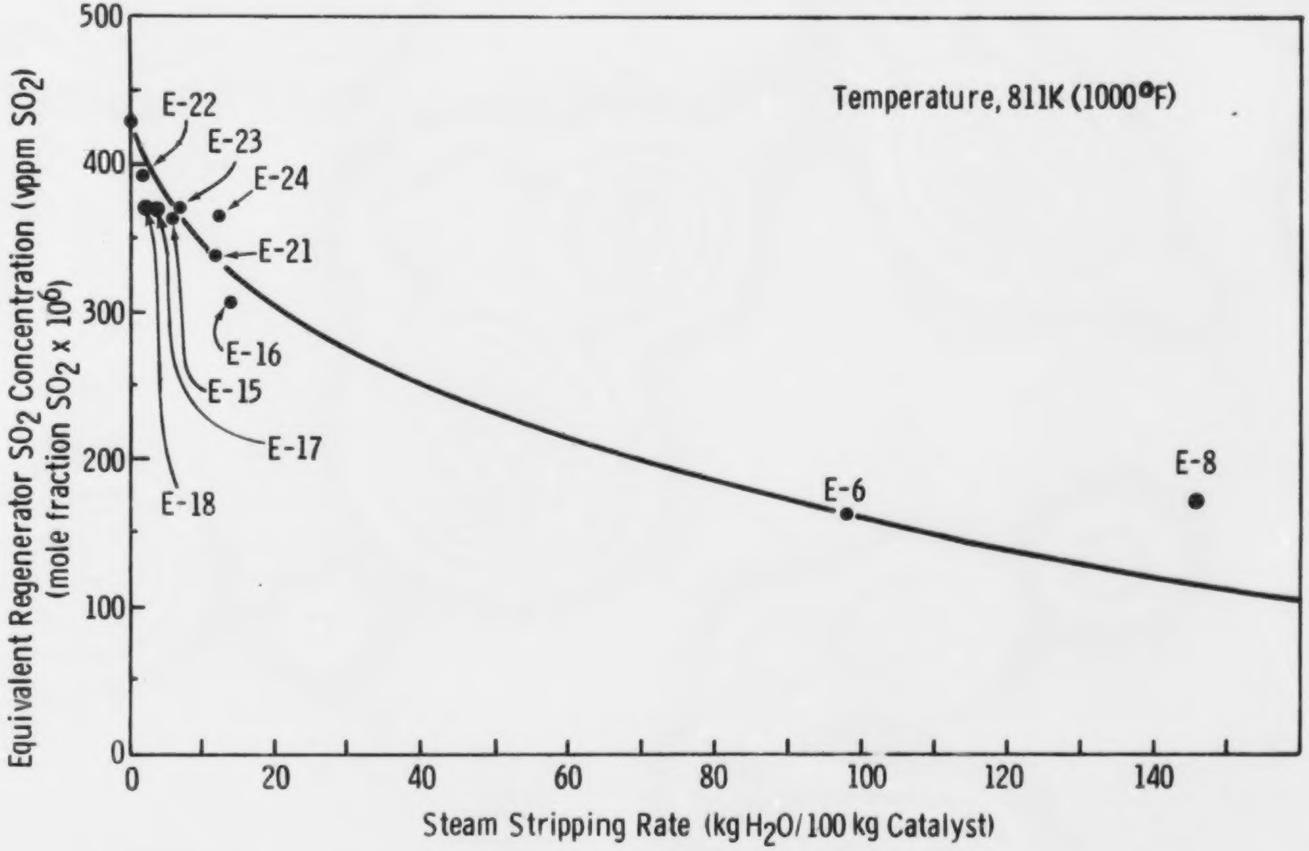

Figure 11. Results of steam stripping experiments on E-Series catalyst with motionless mixer in stripper

Table 6. RESULTS OF STEAM STRIPPING EXPERIMENTS

E-Series Catalyst, without Motionless Mixer

	Experiment					
	E-26	E-28	E-31	E-33	E-36	E-37
Catalyst bed temperature, (K)	811	811	811	811	811	811
(°F)	1000	1000	1000	1000	1000	1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	8.96	7.9	9.64	185	9.20	33.8
Stripper pressure, (Pa)	$1.08\text{x}10^5$	$1.08\text{x}10^5$	$1.08\text{x}10^5$	$1.08\text{x}10^5$	$1.08\text{x}10^5$	$1.08\text{x}10^5$
(psig)	1.0	1.0	1.0	1.0	1.0	1.0
Steam-catalyst contact time, (s)	0.122	0.326	0.0713	0.0741	0.0748	0.0813
Superficial steam-catalyst contact time, T_S, (s)	1.084	3.075	0.674	0.702	0.707	0.769
Catalyst-steam exposure time, (s)	270	675	180	3600	180	720
Stripper superficial velocity, (m/s)	0.341	0.121	0.482	0.445	0.451	0.445
(ft/s)	1.12	0.396	1.58	1.46	1.48	1.46
O_2 content of superheater feedwater, (kg/m^3)	$<2.0\text{x}10^{-5}$	$<2.0\text{x}10^{-5}$	$<2.0\text{x}10^{-5}$	$<2.0\text{x}10^{-5}$	$<2.0\text{x}10^{-5}$	$<2.0\text{x}10^{-5}$
Sulfur removal (% by wt)	29.8	30.6	9.04	23.1	13.3	25.0
Residual equivalent regenerator SO_2 concentration, S_O, (mole fraction)	$3.00\text{x}10^{-4}$	$2.97\text{x}10^{-4}$	$3.89\text{x}10^{-4}$	$3.29\text{x}10^{-4}$	$3.71\text{x}10^{-4}$	$3.21\text{x}10^{-4}$
(vppm)	300	297	389	329	371	321

Remarks: Equivalent regenerator SO_2 concentration before stripping, $4.28\text{x}10^{-4}$ mole fraction on SO_2
428 vppm SO_2 (S_1)

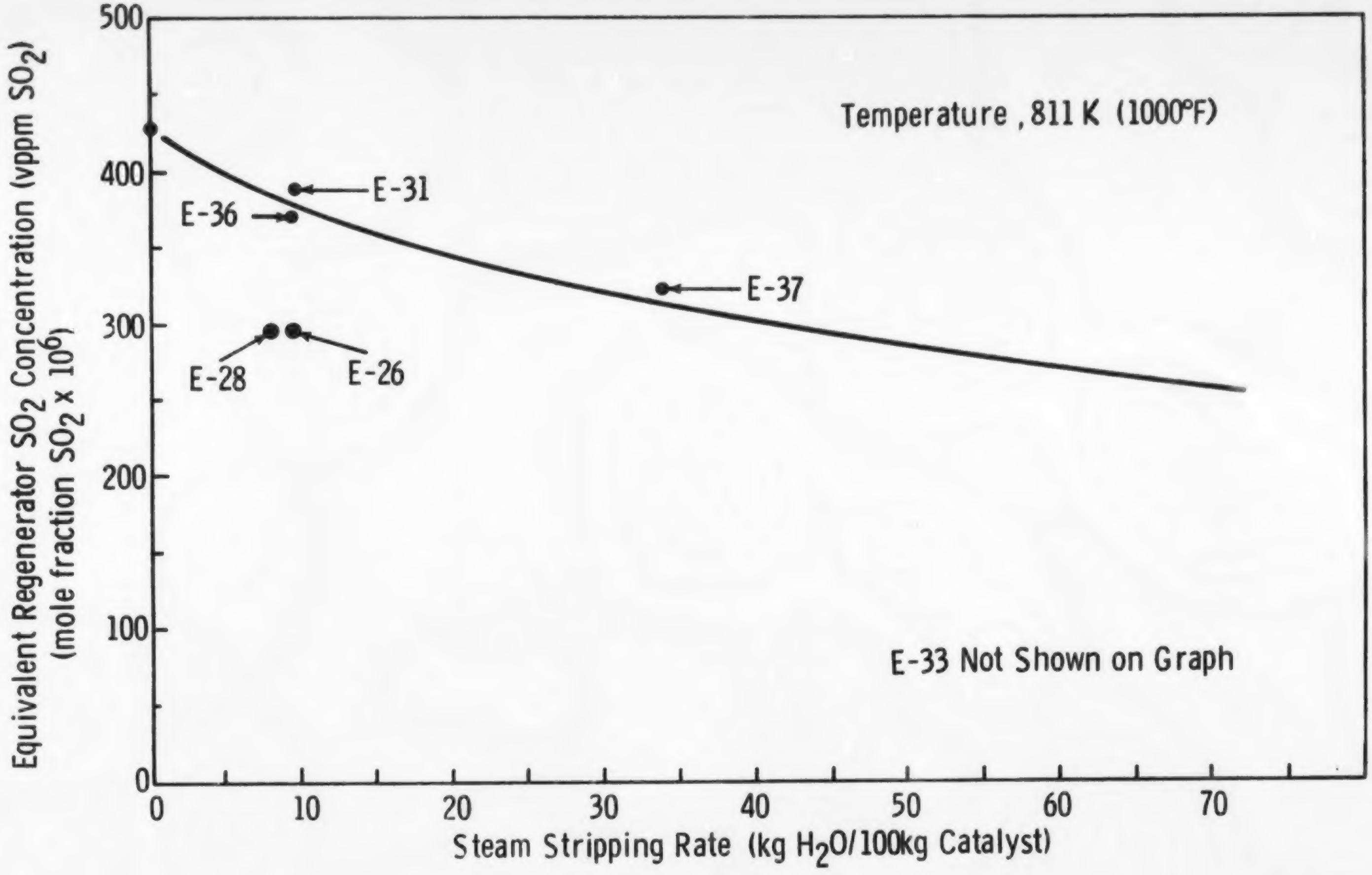

Figure 12. Results of steam stripping experiments on E-Series catalyst without motionless mixer used in the stripper

Table 7. RESULTS OF STEAM STRIPPING EXPERIMENTS

F-Series Catalyst, with Motionless Mixer

	Experiment				
	F-9	F-10	F-11	F-12	F-13
Catalyst bed temperature, (K) (°F)	811 1000	811 1000	811 1000	811 1000	811 1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	5.54	10.5	2.17	6.99	7.41
Stripper pressure, (Pa) (psig)	1.08×10^5 1.0	1.08×10^5 1.0	1.08×10^5 1.0	2.39×10^5 20.0	2.39×10^5 20.0
Steam-catalyst contact time, (s)	0.0807	0.100	0.0787	0.145	0.135
Superficial steam-catalyst contact time, T_S, (s)	0.782	0.927	0.748	1.37	1.29
Catalyst-steam exposure time, (s)	120	270	45	120	120
Stripper superficial velocity, (m/s) (ft/s)	 0.433 1.42	 0.369 1.21	 0.445 1.46	 0.241 0.791	 0.258 0.847
O_2 content of superheater feedwater, (kg/m^3)	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$
Sulfur removal (% by wt)	10.6	16.80	2.69	15.1	10.3
Residual equivalent regenerator SO_2 concentration, S_O, (mole fraction) (vppm)	 4.38×10^{-4} 438	 4.08×10^{-4} 408	 4.77×10^{-4} 477	 4.16×10^{-4} 416	 4.40×10^{-4} 440

Remarks: Equivalent regenerator SO_2 concentration before stripping, 4.90×10^{-4} mole fraction on SO_2
490 vppm SO_2 (S_1)

Table 7 continued. RESULTS OF STEAM STRIPPING EXPERIMENTS
F-Series Catalyst, with Motionless Mixer

	Experiment				
	F-14	F-15	F-17	F-18	F-22
Catalyst bed temperature, (K) (°F)	811 1000	811 1000	811 1000	811 1000	811 1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	11.1	9.57	9.87	9.07	8.20
Stripper pressure, (Pa) (psig)	$2.39x10^5$ 20.0	$2.39x10^5$ 20.0	$3.08x10^5$ 30.0	$3.08x10^5$ 30.0	$3.08x10^5$ 30.0
Steam-catalyst contact time, (s)	0.187	0.166	0.194	0.144	0.272
Superficial steam-catalyst contact time, T_s, (s)	1.72	1.50	1.88	1.36	2.63
Catalyst-steam exposure time, (s)	240	180	180	120	210
Stripper superficial velocity, (m/s) (ft/s)	 0.250 0.819	 0.250 0.819	 0.203 0.665	 0.288 0.946	 0.144 0.474
O_2 content of superheater feedwater, (kg/m^3)	$<2.0x10^{-5}$	$<2.0x10^{-5}$	$<2.0x10^{-5}$	$<2.0x10^{-5}$	$<2.0x10^{-5}$
Sulfur removal (% by wt)	30.1	23.0	14.8	40.8	35.5
Residual equivalent regenerator SO_2 concentration, S_O, (mole fraction) (vppm)	$3.42x10^{-4}$	$3.77x10^{-4}$	$4.17x10^{-4}$	$2.90x10^{-4}$	$3.16x10^{-4}$

Remarks: Equivalent regenerator SO_2 concentration before stripping, $4.90x10^{-4}$ mole fraction SO_2
490 vppm SO_2 (S_1)

Table 7 continued. RESULTS OF STEAM STRIPPING EXPERIMENTS

F-Series Catalyst, with Motionless Mixer

	Experiment				
	F-25	F-27	F-29	F-30	F-31
Catalyst bed temperature, (K) (°F)	811 1000	811 1000	811 1000	811 1000	811 1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	11.9	2.3	192	192	186
Stripper pressure, (Pa) (psig)	3.08×10^5 30.0	1.08×10^5 1.0	1.08×10^5 1.0	2.39×10^5 20.0	3.43×10^5 35.0
Steam-catalyst contact time, (s)	0.255	0.360	0.0714	0.160	0.226
Superficial steam-catalyst contact time, T_S, (s)	2.41	3.29	0.677	1.50	2.14
Catalyst-steam exposure time, (s)	280	210	3600	3600	3600
Stripper superficial velocity, (m/s) (ft/s)	0.163 0.534	0.116 0.379	0.460 1.51	0.194 0.638	0.140 0.460
O_2 content of superheater feedwater, (kg/m^3)	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$
Sulfur removal (% by wt)	47.7	13.8	49.9	49.4	37.3
Residual equivalent regenerator SO_2 concentration, S_O, (mole fraction) (vppm)	2.56×10^{-4} 256	4.22×10^{-4} 422	2.45×10^{-4} 245	2.48×10^{-4} 248	3.07×10^{-4} 307

Remarks: Equivalent regenerator SO_2 concentration before stripping, 4.90×10^{-4} mole fraction SO_2
490 vppm SO_2 (S_1)

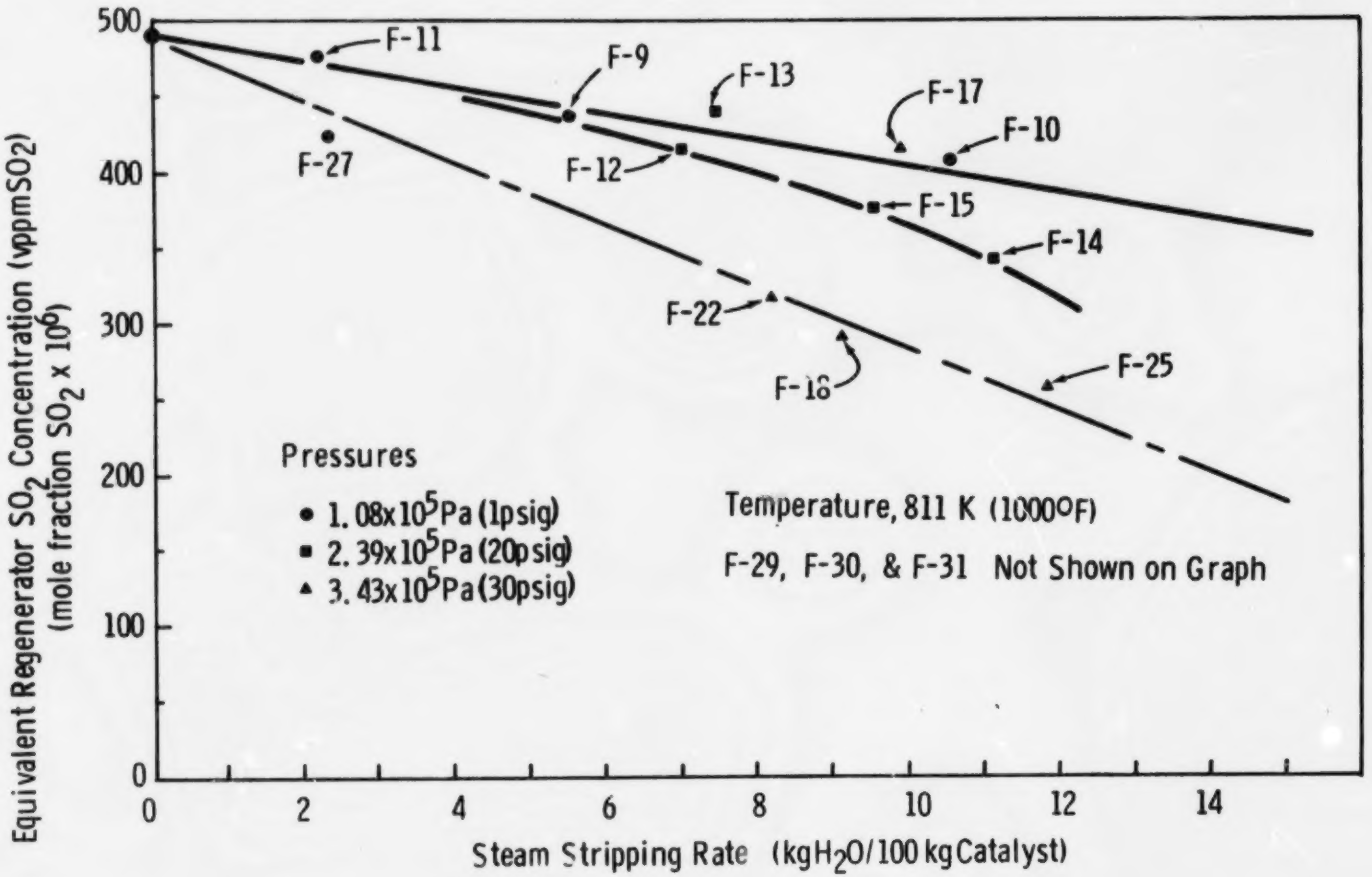

Figure 13. Results of steam stripping experiments on F-Series catalyst

Table 8. RESULTS OF STEAM STRIPPING EXPERIMENTS

H-Series Catalyst, without Motionless Mixer

	Experiment					
	H-5	H-6	H-7	H-8	H-11	H-12
Catalyst bed temperature, (K) (°F)	811 1000	811 1000	811 1000	811 1000	811 1000	811 1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	4.01	7.68	12.9	32.1	11.5	4.37
Stripper pressure, (Pa) (psig)	1.08×10^5 1.0	1.08×10^5 1.0	1.08×10^5 1.0	1.08×10^5 1.0	1.08×10^5 1.0	2.39×10^5 20.
Steam-catalyst contact time, (s)	0.0571	0.0884	0.107	0.0858	0.0800	0.116
Superficial steam-catalyst contact time, T_s, (s)	0.539	0.847	1.01	0.810	0.755	1.10
Catalyst-steam exposure time, (s)	60	180	360	720	2400	60
Stripper superficial velocity, (m/s) (ft/s)	0.689 2.26	0.454 1.49	0.384 1.26	0.475 1.56	0.445 1.46	0.314 1.03
O_2 content of superheater feedwater, (kg/m³)	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$
Sulfur removal (% by wt)	28.2	41.1	44.0	42.9	70.0	54.5
Residual equivalent regenerator SO_2 concentration, S_0, (mole fraction) (vppm)	1.75×10^{-4} 175	1.43×10^{-4} 143	1.36×10^{-4} 136	1.39×10^{-4} 139	0.73×10^{-4} 73	1.10×10^{-4} 110

Remarks: Equivalent regenerator SO_2 concentration before stripping, 2.43×10^{-4} mole fraction SO_2
243 vppm SO_2 (S_1)

Table 8 continued. RESULTS OF STEAM STRIPPING EXPERIMENTS

H-Series Catalyst, without Motionless Mixer

	Experiment					
	H-13	H-14	H-15	H-16	H-17	H-18
Catalyst bed temperature, (K)	811	811	811	811	811	811
(°F)	1000	1000	1000	1000	1000	1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	5.38	8.95	17.1	31.6	64.0	185
Stripper pressure, (Pa)	2.39×10^5	2.39×10^5	2.39×10^5	2.39×10^5	2.39×10^5	2.39×10^5
(psig)	20.0	20.0	20.0	20.0	20.0	20.0
Steam-catalyst contact time, (s)	0.189	0.169	0.177	0.192	0.195	0.164
Superficial steam-catalyst contact time, T_S, (s)	1.78	1.60	1.68	1.82	1.84	1.55
Catalyst-steam exposure time, (s)	120	180	360	720	1500	3600
Stripper superficial velocity, (m/s)	0.204	0.204	0.202	0.185	0.187	0.201
(ft/s)	0.669	0.670	0.662	0.607	0.613	0.659
O_2 content of superheater feedwater, (kg/m^3)	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$
Sulfur removal (% by wt)	43.9	47.8	40.6	51.1	60.8	65.5
Residual equivalent regenerator SO_2 concentration, S_O, (mole fraction)	1.36×10^{-4}	1.27×10^{-4}	1.44×10^{-4}	1.19×10^{-4}	0.95×10^{-4}	0.84×10^{-4}
(vppm)	136	127	144	119	95	84

Remarks: Equivalent regenerator SO_2 concentration before stripping, 2.43×10^{-4} mole fraction SO_2
243 vppm SO_2 (S_1)

Table 8 continued. RESULTS OF STEAM STRIPPING EXPERIMENTS

H-Series Catalyst, without Motionless Mixer

	Experiment						
	H-19	H-21	H-22	H-23	H-24	H-25	H-26
Catalyst bed temperature, (K)	811	811	811	811	811	811	811
(°F)	1000	1000	1000	1000	1000	1000	1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	4.67	164	6.89	9.78	15.6	41.8	65.6
Stripper pressure, (Pa)	3.43×10^5	3.43×10^5	3.43×10^5	3.43×10^5	3.43×10^5	3.43×10^5	3.43×10^5
(psig)	35.0	35.0	35.0	35.0	35.0	35.0	35.0
Steam-catalyst contact time, (s)	0.151	0.256	0.204	0.216	0.271	0.201	0.268
Superficial steam-catalyst contact time, T_s, (s)	1.42	2.42	1.92	2.03	2.54	1.90	2.53
Catalyst-steam exposure time, (s)	60	3600	120	180	360	720	1500
Stripper superficial velocity, (m/s)	0.225	0.136	0.163	0.148	0.134	0.159	0.134
(ft/s)	0.737	0.445	0.535	0.485	0.439	0.521	0.439
O_2 content of superheater feedwater, (kg/m^3)	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	2.0×10^{-5}	2.0×10^{-5}	2.0×10^{-5}	2.0×10^{-5}
Sulfur removal (% by wt)	32.6	73.9	39.2	42.2	56.3	59.1	55.0
Residual equivalent regenerator SO_2 concentration, S_O, (mole fraction)	1.64×10^{-4}	0.63×10^{-4}	1.48×10^{-4}	126×10^{-4}	1.06×10^{-4}	0.99×10^{-4}	1.80×10^{-4}
(vppm)	164	63	148	126	106	99	109

Remarks: Equivalent regenerator SO_2 concentration before stripping, 2.43×10^{-4} mole fraction SO_2
vppm SO_2 (S_1)

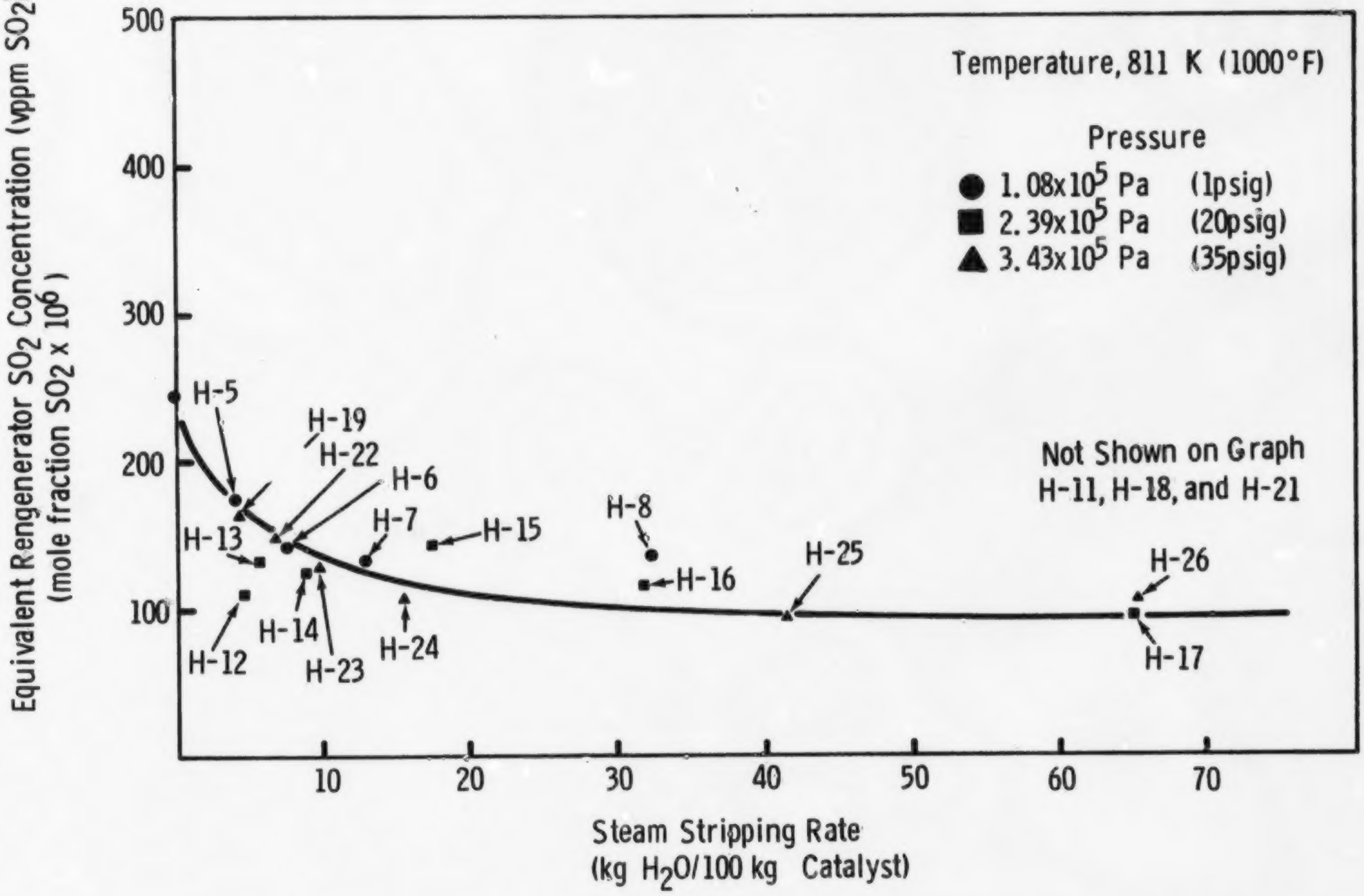

Figure 14. Results of steam stripping experiments on H-Series catalyst

Table 9. RESULTS OF STEAM STRIPPING EXPERIMENTS

H-Series Catalyst, without Motionless Mixer

	Experiment					
	H-27	H-28	H-29	H-30	H-31	H-32
Catalyst bed temperature, (K)	755	755	755	755	755	755
(°F)	900	900	900	900	900	900
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	14.6	72.0	179	18.2	10.5	8.81
Stripper pressure, (Pa)	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5
(psig)	1.0	1.0	1.0	1.0	1.0	1.0
Steam-catalyst contact time, (s)	0.10	0.0851	0.0825	0.8080	.0705	0.0557
Superficial steam-catalyst contact time, T_S, (s)	0.953	0.803	0.777	0.765	0.663	0.525
Catalyst-steam exposure time, (s)	360	1500	3600	360	180	120
Stripper superficial velocity, (m/s)	0.347	0.387	0.415	0.418	0.482	0.631
(ft/s)	1.14	1.27	1.36	1.37	1.58	2.07
O_2 content of superheater feedwater, (kg/m³)	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$
Sulfur removal (% by wt)	57.6	57.6	65	54.2	58.2	26.0
Residual equivalent regenerator SO_2 concentration, S_0, (mole fraction)	1.03×10^{-4}	1.03×10^{-4}	0.85×10^{-4}	1.11×10^{-4}	1.02×10^{-4}	1.80×10^{-4}
(vppm)	103	103	85	111	102	180

Remarks: Equivalent regenerator SO_2 concentration before stripping, 2.43×10^{-4} mole fraction SO_2
243 vppm SO_2 (S_1)

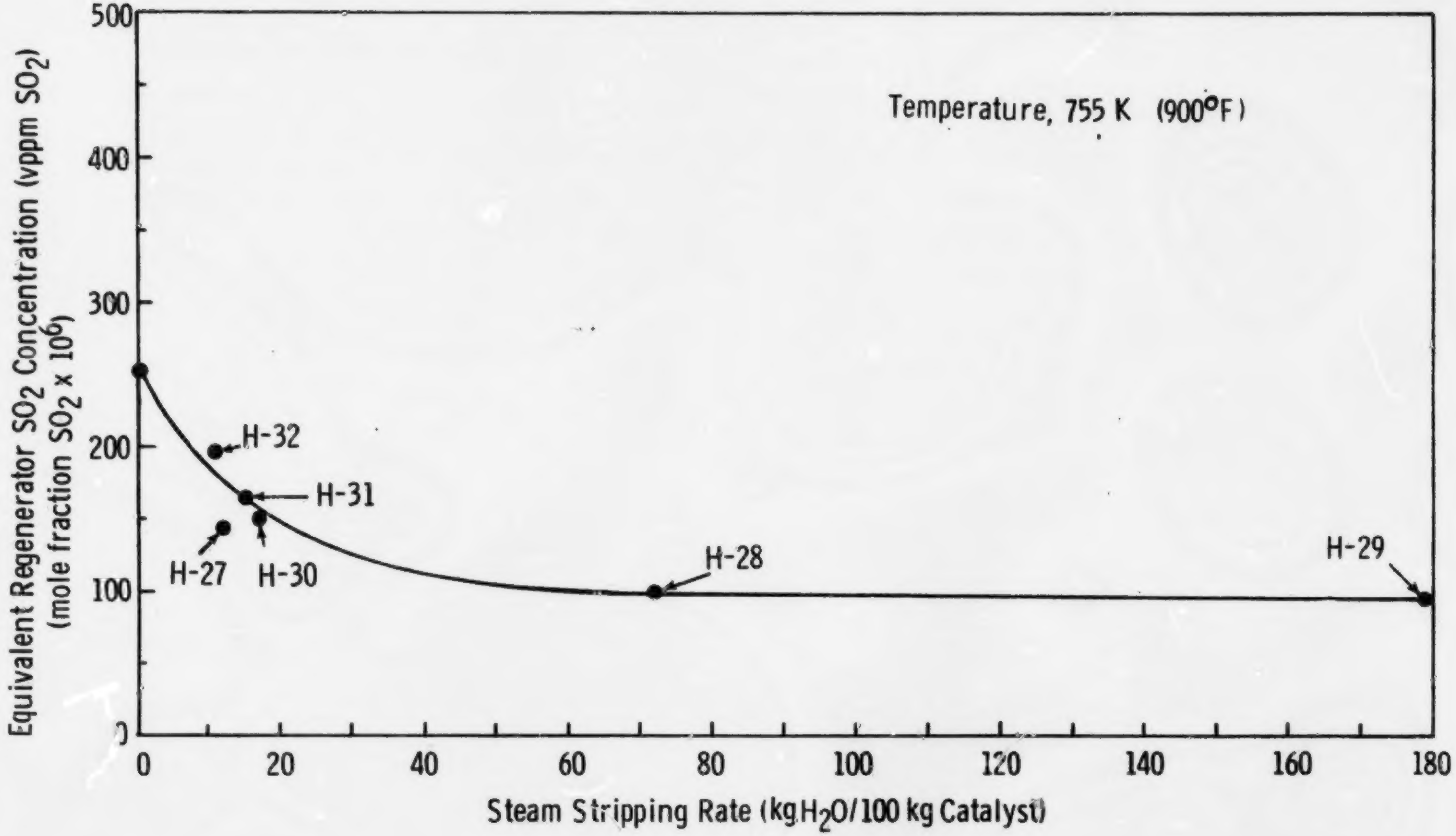

Figure 15. Results of steam stripping experiments of H-Series catalyst

Table 10. RESULTS OF STEAM STRIPPING EXPERIMENTS

I-Series Catalyst, without Motionless Mixer

	Experiment				
	I-5	I-8	I-9	I-10	I-11
Catalyst bed temperature, (K)	811	811	811	811	811
(°F)	1000	1000	1000	1000	1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	4.19	29.1	53.4	194	13.4
Stripper pressure, (Pa)	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5
(psig)	1.0	1.0	1.0	1.0	1.0
Steam-catalyst contact time, (s)	0.0546	0.0631	0.690	0.0710	0.0668
Superficial steam-catalyst contact time, T[illegible], (s)	0.517	0.596	0.648	0.669	0.648
Catalyst-steam exposure time, (s)	60	480	960	3600	240
Stripper superficial velocity, (m/s)	0.524	0.475	0.433	0.436	0.451
(ft/s)	1.72	1.56	1.42	1.43	1.48
O_2 content of superheater feedwater, (kg/m^3)	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	<2.0x10-5	<2.0x10-5
Sulfur removal (% by wt)	22.58	37.27	41.48	50.64	31.52
Residual equivalent regenerator SO_2 concentration, S_0, (mole fraction)	4.77×10^{-4}	3.86×10^{-4}	3.61×10^{-4}	3.04×10^{-4}	4.22×10^{-4}
(vppm)	477	386	361	304	422

Remarks: Equivalent regenerator SO_2 concentration before stripping, 6.16×10^{-4} mole fraction on SO_2
616 vppm SO_2 (S_1)

Table 10 continued. RESULTS OF STEAM STRIPPING EXPERIMENTS

I-Series Catalyst, without Motionless Mixer

	Experiment				
	I-12	I-13	I-14	I-15	I-16
Catalyst bed temperature, (K)	811	811	811	811	811
(°F)	1000	1000	1000	1000	1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	3.99	31.6	97.1	182	49.1
Stripper pressure, (Pa)	2.39×10^5	2.39×10^5	2.39×10^5	2.39×10^5	2.39×10^5
(psig)	20	20	20	20	20
Steam-catalyst contact time, (s)	0.254	0.128	0.156	0.167	0.155
Superficial steam-catalyst contact time, T_s, (s)	2.40	1.21	1.48	1.58	1.46
Catalyst-steam exposure time, (s)	120	480	1800	3600	900
Stripper superficial velocity, (m/s)	0.120	0.232	0.191	0.215	0.201
(ft/s)	0.395	0.761	0.628	0.704	0.659
O_2 content of superheater feedwater, (kg/m³)	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$	$<2.0 \times 10^{-5}$
Sulfur removal (% by wt)	25.31	35.02	50.67	55.45	51.31
Residual equivalent regenerator SO_2 concentration, S_0, (mole fraction)	4.60×10^{-4}	4.00×10^{-4}	3.04×10^{-4}	2.74×10^{-4}	3.00×10^{-4}
(vppm)	460	400	304	274	300

Remarks: Equivalent regenerator SO_2 concentration before stripping, 6.16×10^{-4} mole fraction on SO_2
616 vppm SO_2 (S_1)

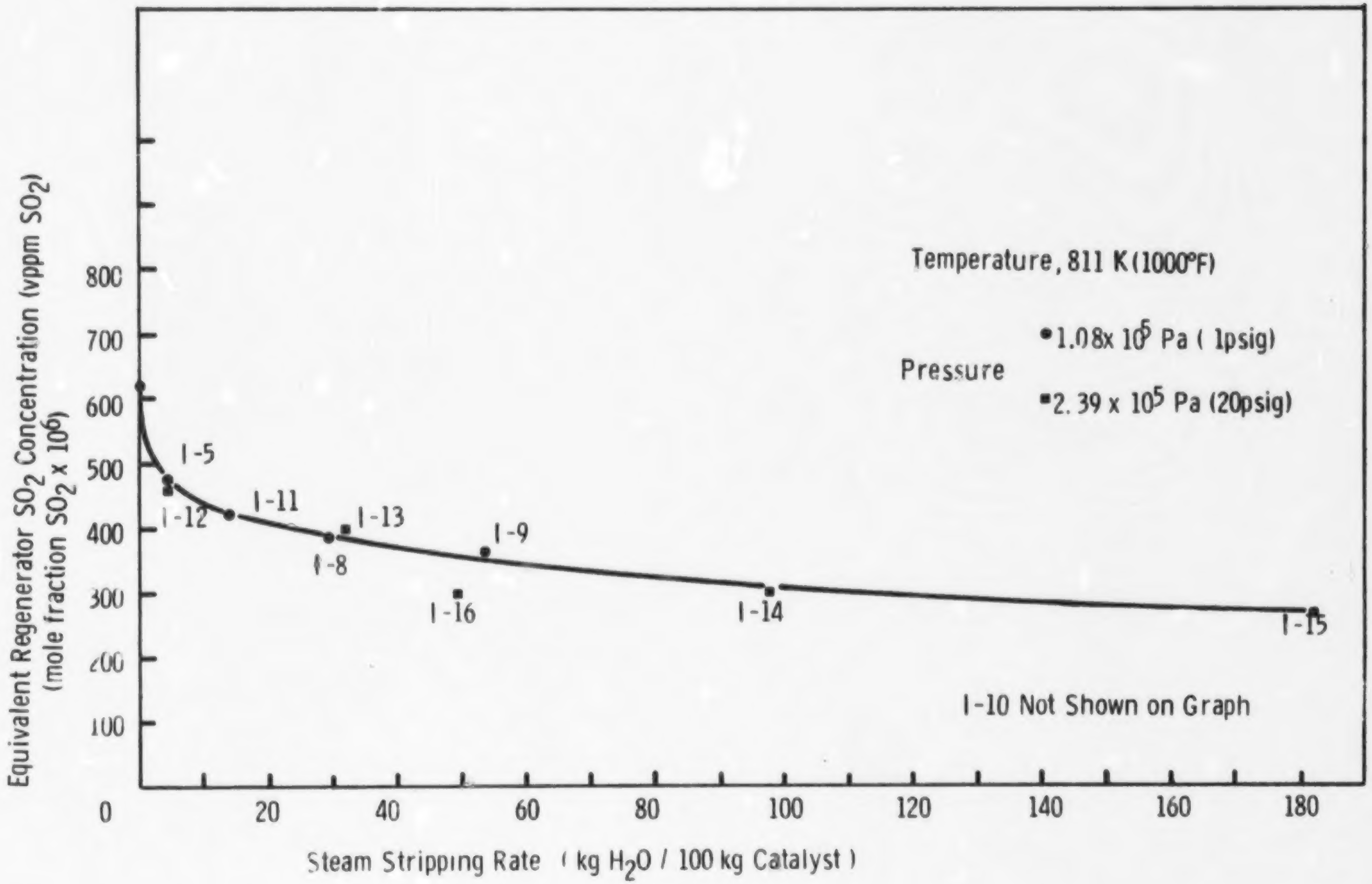

Figure 16. Results of steam stripping experiments on I-Series catalyst

Table 11. RESULTS OF EXPERIMENTS PERFORMED TO DETERMINE THE EFFECT OF STEAM STRIPPING ON COKE VOLATILITY

B-Series Catalyst

	Experiment			
	B-3	B-5	B-7	B-8
Catalyst bed temperature, (K) (°F)	811 1000	811 1000	811 1000	811 1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	5.31	9.14	17.6	39.8
Stripper pressure, (Pa) (psig)	1.08×10^5 1.0	2.39×10^5 20.0	2.39×10^5 20.0	2.39×10^5 20.0
Steam-catalyst contact time, (s)	0.123	0.166	0.145	0.156
Superficial steam-catalyst contact time, T_s, (s)	1.23	1.57	1.36	1.48
Catalyst-steam exposure time, (s)	180	180	300	738
Stripper superficial velocity, (m/s) (ft/s)	0.276 0.907	0.197 0.646	0.155 0.509	0.197 0.646
Carbon content of stripper condensate (kg/m^3)	1.016	0.714	0.518	0.168
Carbon removal (% by weight of coke)	0.817	0.989	1.45	1.01

Table 11 continued. RESULTS OF EXPERIMENTS PERFORMED TO DETERMINE THE EFFECT OF STEAM STRIPPING ON COKE VOLATILITY

B-Series Catalyst

	Experiment			
	B-9	B-11	B-12	B-13
Catalyst bed temperature, (K)	811	811	811	811
(°F)	1000	1000	1000	1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	9.01	21.9	99.6	1372
Stripper pressure, (Pa)	3.08×10^5	3.43×10^5	3.43×10^5	3.43×10^5
(psig)	30.0	35.0	35.0	35.0
Steam-catalyst contact time, (s)	0.212	0.192	0.212	0.0164
Superficial steam-catalyst contact time, T_s, (s)	2.05	1.81	2.00	0.155
Catalyst-steam exposure time, (s)	180	360	1800	1920
Stripper superficial velocity, (m/s)	0.134	0.158	0.141	0.140
(ft/s)	0.438	0.517	0.460	0.458
Carbon content of stripper condensate (kg/m^3)	0.739	0.215	0.0721	0.0340
Carbon removal (% by weight of coke)	1.01	0.713	1.09	7.07

Table 12. RESULTS OF EXPERIMENTS PERFORMED TO DETERMINE THE EFFECT OF STEAM STRIPPING ON COKE VOLATILITY

C-Series Catalyst

	Experiment						
	C-6	C-7	C-10	C-11	C-12	C-13	C-14
Catalyst bed temperature, (K) (°F)	811 1000	811 1000	811 1000	811 1000	811 1000	811 1000	811 1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	7.31	14.1	16.0	5.98	37.2	109	29.5
Stripper pressure, (Pa) (psig)	1.08×10^5 1.0	1.08×10^5 1.0	2.39×10^5 20.0	2.39×10^5 20.0	2.39×10^5 20.0	3.43×10^5 35.0	3.43×10^5 35.0
Steam-catalyst contact time, (s)	0.0939	0.0971	0.191	0.134	0.164	0.258	0.286
Superficial steam-catalyst contact time, T_s, (s)	0.888	0.924	1.80	1.27	1.54	2.43	2.69
Catalyst-steam exposure time, (s)	180	360	360	95	720	2400	720
Stripper superficial velocity, (m/s) (ft/s)	0.396 1.30	0.408 1.34	0.212 0.695	0.258 0.845	0.199 0.652	0.156 0.475	0.135 0.444
Carbon content of stripper condensate (kg/m^3)	1.498	0.325	0.427	0.863	0.217	0.008	0.797
Carbon removal (% by weight of coke)	0.911	0.793	1.18	0.896	1.40	1.52	4.09

Table 12 continued. RESULTS OF EXPERIMENTS PERFORMED TO DETERMINE THE EFFECT OF STEAM STRIPPING ON COKE VOLATILITY

C-Series Catalyst

	Experiment					
	C-15	C-16*	C-18*	C-20	C-21	C-24
Catalyst bed temperature, (K) (°F)	811 1000	811 1000	811 1000	811 1000	811 1000	811 1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	7.59	115	109	738	631	177
Stripper pressure, (Pa) (psig)	3.43×10^5 35.0	1.08×10^5 1.0	3.43×10^5 35.0	2.39×10^5 20.0	3.43×10^5 35.0	1.08×10^5 1.0
Steam-catalyst contact time, (s)	0.185	0.0799	0.258	0.0404	0.0657	0.0730
Superficial steam-catalyst contact time, T_S, (s)	1.75	0.756	2.43	0.389	0.631	0.733
Catalyst-steam exposure time, (s)	120	2400	2400	3600	3600	3600
Stripper superficial velocity, (m/s) (ft/s)	0.215 0.704	0.445 1.46	0.154 0.505	0.205 0.673	0.143 0.469	0.460 1.51
Carbon content of stripper condensate (kg/m^3)	1.524	1.355	1.899	0.0214	0.0406	0.0464
Carbon removal (% by weight of coke)	2.01	27.0	36.0	2.75	4.45	1.43

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* Samples contaminated with acetone from bottle

Table 13. RESULTS OF EXPERIMENTS PERFORMED TO DETERMINE THE EFFECT OF STEAM STRIPPING ON COKE VOLATILITY

E-Series Catalyst

	Experiment				
	E-31	E-33	E-35	E-36	E-37
Catalyst bed temperature, (K) (°F)	811 1000	811 1000	811 1000	811 1000	811 1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	9.64	185	70.5	9.2	33.8
Stripper pressure, (Pa) (psig)	1.08×10^5 1.0	1.08×10^5 1.0	1.08×10^5 1.0	1.08×10^5 1.0	1.08×10^5 1.0
Steam-catalyst contact time, (s)	0.713	0.0741	0.0814	0.748	0.0813
Superficial steam-catalyst contact time, T_S, (s)	0.674	0.702	0.768	0.707	0.769
Catalyst-steam exposure time, (s)	180	3600	1500	180	720
Stripper superficial velocity, (m/s) (ft/s)	0.482 1.58	0.445 1.46	0.485 1.49	0.482 1.48	0.445 1.46
Carbon content of stripper condensate (kg/m^3)	0.246	0.0532	0.0554	3.026	0.531
Carbon removal (% by weight of coke)	0.232	0.966	0.383	2.73	1.76

Table 14. RESULTS OF EXPERIMENTS PERFORMED TO DETERMINE THE EFFECT OF STEAM STRIPPING ON COKE VOLATILITY

F-Series Catalyst

	Experiment		
	F-29	F-30	F-31
Catalyst bed temperature, (K)	811	811	811
(°F)	1000	1000	1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	192	192	186
Stripper pressure, (Pa)	1.08×10^5	2.39×10^5	3.43×10^5
(psig)	1.0	20.0	35.0
Steam-catalyst contact time, (s)	0.0714	0.160	0.226
Superficial steam-catalyst contact time, T_s, (s)	0.677	1.50	2.14
Catalyst-steam exposure time, (s)	3600	3600	3600
Stripper superficial velocity, (m/s)	0.460	0.194	0.140
(ft/s)	1.51	0.638	0.460
Carbon content of stripper condensate (kg/m^3)	0.126	0.0879	0.110
Carbon removal (% by weight of coke)	2.83	1.97	2.40

Table 15. RESULTS OF EXPERIMENTS PERFORMED TO DETERMINE THE EFFECT OF STEAM STRIPPING ON COKE VOLATILITY

H-Series Catalyst

	Experiment						
	H-5	H-6	H-7	H-8	H-9	H-11	H-12
Catalyst bed temperature, (K)	811	811	811	811	811	811	811
(°F)	1000	1000	1000	1000	1000	1000	1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	4.01	7.68	12.9	32.1	69.0	115	4.37
Stripper pressure, (Pa)	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5	2.39×10^5
(psig)	1.0	1.0	1.0	1.0	1.0	1.0	20.0
Steam-catalyst contact time, (s)	0.0571	0.0884	0.107	0.0858	0.0831	0.0800	0.116
Superficial steam-catalyst contact time, T_s, (s)	0.539	0.847	1.0	0.810	0.784	0.755	1.10
Catalyst-steam exposure time, (s)	60	180	360	720	1500	2400	60
Stripper superficial velocity, (m/s)	0.689	0.454	0.384	0.475	0.463	0.445	0.314
(ft/s)	2.26	1.49	1.26	1.56	1.52	1.46	1.03
Carbon content of stripper condensate (kg/m^3)	1.419	1.541	1.957	0.289	0.165	0.0300	1.987
Carbon removal (% by weight of coke)	0.437	0.907	1.94	0.712	0.871	0.261	0.255

Table 15 continued. RESULTS OF EXPERIMENT PERFORMED TO DETERMINE THE EFFECT OF STEAM STRIPPING ON COKE VOLATILITY

H-Series Catalyst

	Experiment						
	H-13	H-14	H-15	H-16	H-17	H-18	H-19
Catalyst bed temperature, (K)	811	811	811	811	811	811	811
(°F)	1000	1000	1000	1000	1000	1000	1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	5.38	8.95	17.1	31.6	65.0	185	4.67
Stripper pressure, (Pa)	2.39×10^5	2.39×10^5	2.39×10^5	2.39×10^5	2.39×10^5	2.39×10^5	3.43×10^5
(psig)	20.0	20.0	20.0	20.0	20.0	20.0	35.0
Steam-catalyst contact time, (s)	0.189	0.169	0.177	0.192	0.195	0.164	0.151
Superficial steam-catalyst contact time, T_S, (s)	1.78	1.60	1.68	1.82	1.84	1.55	1.42
Catalyst-steam exposure time, (s)	120	180	360	720	1500	3600	60
Stripper superficial velocity, (m/s)	0.204	0.204	0.202	0.185	0.187	0.201	0.225
(ft/s)	0.669	0.670	0.662	0.607	0.613	0.659	0.737
Carbon content of stripper condensate (kg/m^3)	0.379	1.176	0.332	0.0581	0.173	0.0097	1.742
Carbon removal (% by weight of coke)	0.157	0.808	0.435	0.141	0.864	0.803	0.624

Table 15 continued. RESULTS OF EXPERIMENTS PERFORMED TO DETERMINE THE EFFECT OF STEAM STRIPPING ON COKE VOLATILITY

H-Series Catalyst

	Experiment					
	H-21	H-22	H-23	H-24	H-25	H-26
Catalyst bed temperature, (K)	811	811	811	811	811	811
(°F)	1000	1000	1000	1000	1000	1000
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	164	6.89	9.78	15.6	41.8	65.6
Stripper pressure, (Pa)	3.43×10^5	3.43×10^5	3.43×10^5	3.43×10^5	3.43×10^5	3.43×10^5
(psig)	35.0	35.0	35.0	35.0	35.0	35.0
Steam-catalyst contact time, (s)	0.256	0.204	0.216	0.271	0.201	0.268
Superficial steam-catalyst contact time, τ_s, (s)	2.42	1.92	2.03	2.54	1.90	2.53
Catalyst-steam exposure time, (s)	3600	120	180	360	720	1500
Stripper superficial velocity, (m/s)	0.136	0.163	0.148	0.134	0.159	0.134
(ft/s)	0.445	0.535	0.485	0.439	0.521	0.439
Carbon content of stripper condensate (kg/m^3)	0.0545	0.222	0.177	0.446	0.0548	0.208
Carbon removal (% by weight of coke)	0.686	0.117	0.134	0.535	0.176	1.05

Table 16. RESULTS OF EXPERIMENTS PERFORMED TO DETERMINE THE EFFECT OF STEAM STRIPPING ON COKE VOLATILITY

H-Series Catalyst

	Experiment					
	H-27	H-28	H-29	H-30	H-31	H-32
Catalyst bed temperature, (K)	755	755	755	755	755	755
(°F)	900	900	900	900	900	900
Steam stripping rate (SSR), kg H_2O/100 kg catalyst)	14.6	72.0	179	18.2	10.5	8.81
Stripper pressure, (Pa)	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5	1.08×10^5
(psig)	1.0	1.0	1.0	1.0	1.0	1.0
Steam-catalyst contact time, (s)	0.101	0.0851	0.0825	0.0808	0.0705	0.0557
Superficial steam-catalyst contact time, T_S, (s)	0.953	0.803	0.777	0.765	0.663	0.525
Catalyst-steam exposure time, (s)	360	1500	3600	360	180	120
Stripper superficial velocity, (m/s)	0.347	0.387	0.415	0.418	0.482	0.631
(ft/s)	1.14	1.27	1.36	1.37	1.58	2.07
Carbon content of stripper condensate (kg/m^3)	0.627	0.0536	0.0651	0.655	2.292	0.237
Carbon removal (% by weight of coke)	0.701	0.285	0.900	0.911	1.78	0.165

Regardless of the catalyst type, the carbon content of stripper condensate was plotted against the steam stripping on a logarithmic paper, Figure 17, and a rather uniform relationship between the two variables was observed. Regression analysis of the data produced the following equations.

$$\log \text{TOC} = -0.998 \log (\text{SSR}) + 3.909 \qquad (1)$$

$$\log (\text{SSR}) = -1.002 \log \text{TOC} + 3.918 \qquad (2)$$

4.2.4 Volatility of Coke on Spent Catalyst

In order to obtain information on types of compounds that may volatilize from the catalyst at elevated temperatures we have performed the catalyst coke volatility test. The procedure followed to perform this test was described in Section 4.1.6.3. As indicated by the procedure, the test was carried out under vacuum and therefore is not an exact simulation of conditions that would exist in the catalyst stripper where some pressure is present and the coke is exposed to continuous flow of steam. Also, during steam stripping the steam may react with the hydrocarbons and produce other hydrocarbon product mix. However, we feel the test together with other experiments made in this study may provide some information as to the type of compounds that can volatilize during the catalyst steam stripping.

Two catalyst samples, "H" - and "E"-series, were used for this test. The results are summarized in Tables 17 and 18. Other catalyst samples were not tested because the purpose of the test was to demonstrate that the presence of

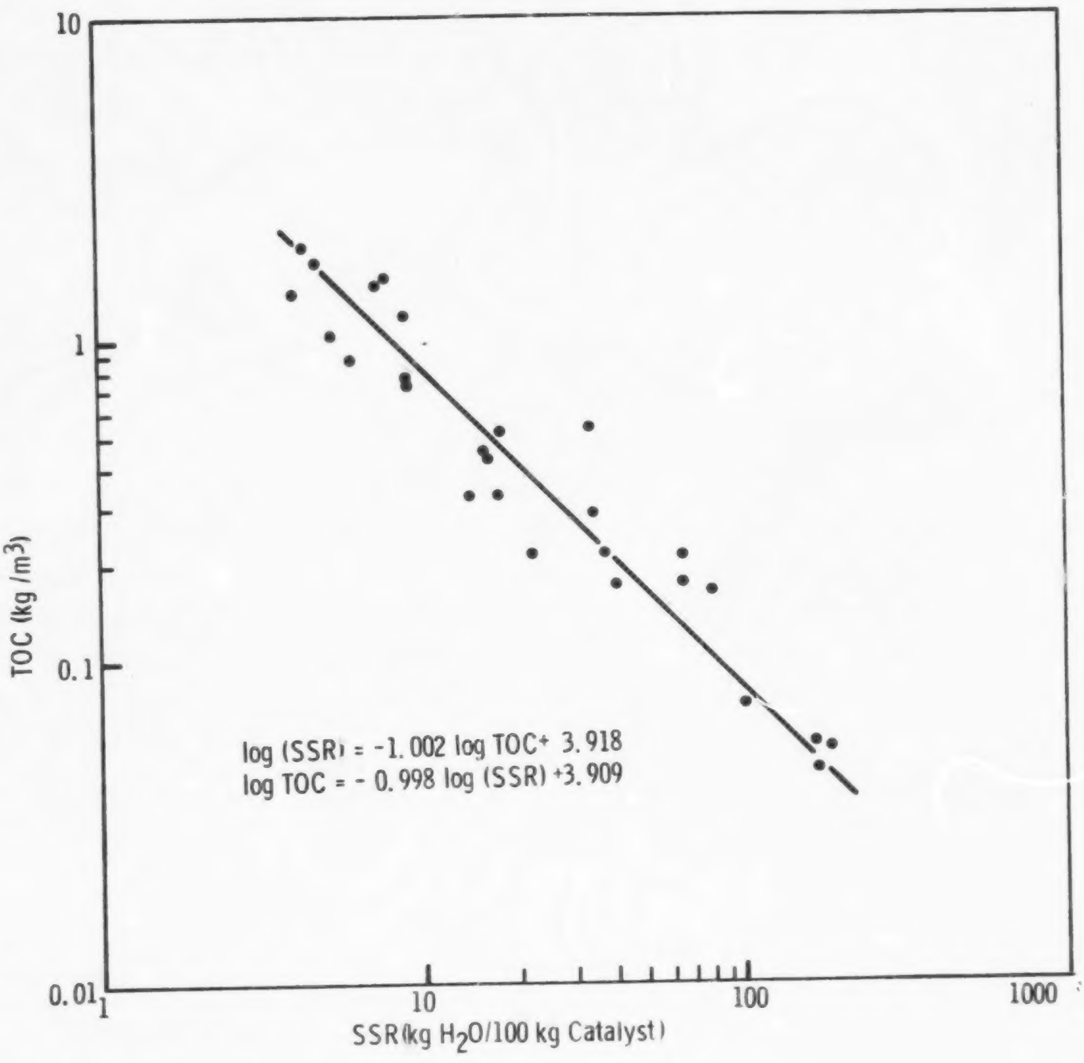

Figure 17. Effect of steam stripping rate upon total organic carbon content of stripper condensate

Table 17. RESULTS OF COKE VOLATILITY EXPERIMENTS ON H-SERIES CATALYST

Determination of Coke Volatility
Not Due to Steam Stripping

Temperature Range	Components[a] (wt. % of catalyst) CH_4	C_3H_6	C_6H_6
293 - 423 K 68 - 302°F	0.001	0.01	ND
423 - 573 K 302 - 572°F	0.002	0.02	0.0092
573 - 673 K 572 - 752°F	0.001	0.01	0.0003
673 - 811 K 752 - 1000°F	0.001	0.01	0.0008
totals[b] (C + H)	0.005	0.05	0.0103
(C)	0.0038	0.043	0.0088

[a] These were the only compounds detected in the gases evolved.

[b] C + H = weight percent of catalyst volatilized as hydrocarbon.

C = weight percent of catalyst volatilized as carbon (obtained by calculating carbon content of hydrocarbon volatilized).

Table 18. RESULTS OF COKE VOLATILITY EXPERIMENTS ON E-SERIES CATALYST

Determination of Coke Volatility Not Due to Steam Stripping

	Components[a] (wt. % of catalyst)			
Temperature Range	CH_4	C_2H_4	C_3H_6	C_4H_8
293 - 373 K 68 - 212°F	0.0004	0.006	0.005	ND
373 - 573 K 212 - 572°F	0.001	0.01	0.01	0.009
573 - 811 K 572 - 1000°F	0.001	0.003	0.004	ND
totals[b] (C + H)	0.0024	0.019	0.019	0.009
(C)	0.0018	0.016	0.016	0.0077

[a] These were the only compounds detected in the gases evolved.

[b] C + H = weight percent of catalyst volatilized as hydrocarbon.
C = weight percent of catalyst volatilized as carbon (obtained by calculating carbon content of hydrocarbon volatilized).

hydrocarbons in the stripper off-gases may not be entirely due to the action of steam upon the coke, but also hydrocarbon volatilization at elevated temperatures of up to 752K (1000°F).

4.2.5 By-Product Formation During Steam Stripping

In order to determine the extent of by-product formation during steam stripping, several tests were made. These tests were performed to identify the types and concentrations of compounds formed during catalyst steam stripping.

The tests were performed following the same procedure as that used for steam stripping experiments. The effluent gas collection system consisted of an ice water sampling train and a Tedlar bag. Thus, no constitutents could leave the system. After the condensation of stripping steam the fluidized bed reactor and the ice water sampling train were flushed with nitrogen and all gases collected in the Tedlar bag. Both the condensate and the contents of the Tedlar bag were analyzed gas chromatographically. In addition, the condensate was analyzed for ammonia. The compounds formed include CH_4, C_2H_4, C_3H_6, COS, CS_2, H_2S, SO_2, and NH_3.

Table 19 presents the actual results of the tests. The operation of the ice water sampling train requires some water present in the train to absorb sulfur dioxide or hydrogen sulfide. Consequently, the results obtained by a direct analysis of the ice water train contents involve some dilution of compounds originally present in the stripping steam. Only compounds detected appear in Table 19.

Table 19. SUMMARY OF BY-PRODUCT FORMATION EXPERIMENTS

Sulfur and Hydrocarbon Compounds

Sample No.	Medium	Compounds Present	Concentration of Compounds, (gas - mole fraction) (liq - kg/m^3)	N_2-free Gas Compositions (dry basis) (vol. %)	Fraction of Feed Sulfur & Carbon in Sample (wt %)
D-4	Gas	CH_4	630×10^{-6}	58.0	0.214 carbon
		C_3H_6	13×10^{-6}	1.2	0.0132 carbon
		H_2S	443×10^{-6}	40.8	64.3 sulfur
		N_2	Balance	--	--
	Liquid	COS	1.5×10^{-3}	--	1.21 sulfur
C-25	Gas	CH_4	1400×10^{-6}	70.3	0.445 carbon
		C_2H_4	99×10^{-6}	4.9	0.063 carbon
		H_2S	492×10^{-6}	24.8	41.2 sulfur
		N_2	Balance	--	--
	Liquid	None	ND	--	--
C-26	Gas	CH_4	2600×10^{-6}	71.6	0.968 carbon
		C_2H_4	273×10^{-6}	7.5	0.203 carbon
		C_3H_6	42×10^{-6}	1.2	--
		H_2S	713×10^{-6}	19.7	67.4 sulfur
		N_2	Balance	--	--
	Liquid	COS	0.54×10^{-3}	--	0.258 sulfur
H-35	Gas	CH_4	137×10^{-6}	51.2	0.0468 carbon
		COS	1.47×10^{-6}	0.3	0.453 sulfur
		H_2S	130×10^{-6}	48.5	40.1 sulfur
		N_2	Balance	--	--
	Liquid	COS	0.188×10^{-3}	--	0.25 sulfur

ND = not detected

Since the amount of stripping steam used in these experiments was known, we have recalculated the results in Table 19 to obtain apparent average concentrations of the detected compounds in stripping steam. The results of these calculations appear in Table 20, 21, and 22.

4.2.6 Effect of Steam Stripping on Catalyst Activity

A serious concern was raised regarding the possible deactivation of FCC catalyst during its extended exposure to steam. Several tests were performed to determine the effect of steam stripping on catalyst activity. The results of these tests appear in Table 23.

Five catalyst samples identified by six-digit numbers and included in Table 23 were analyzed by Davison Chemical Division, W. R. Grace Company, Baltimore, Md. The analyses included chemical analysis, physical analysis, and activity determinations. More specific identification of the five samples follows.

Sample 165445 was an "as-received" catalyst. This sample was neither steam stripped nor regenerated in our catalyst test unit. It was used to establish the basis for comparison of samples received from the refinery and those later exposed to steam stripping experiments.

Sample 165446 was not steam stripped but was regenerated at 1.08 x 10^5 Pa (1 psig) and 866 K (1100°F) for 3600 s (60 minutes) in our catalyst test unit. This sample was analyzed to determine whether or not our catalyst regeneration technique caused any catalyst deactivation.

Table 20. CALCULATED STRIPPER OFF-GAS ANALYSIS C-Series Catalyst (C-26)

Component	Concentration, mole fraction
H_2S	2520×10^{-6}
COS	1.78×10^{-6}
CS_2	N D
CH_4	7200×10^{-6}
C_2H_4	50.5×10^{-6}
C_3H_6	N D
NH_3	445×10^{-6}
H_2O	balance

ND = not detected

Table 21. CALCULATED STRIPPER OFF-GAS ANALYSIS
H-Series Catalyst (H-35)

Component	Concentration, mole fraction
H_2S	116×10^{-6}
COS	2.1×10^{-6}
CS_2	N D
CH_4	35×10^{-6}
C_2H_4	N D
C_3H_6	N D
NH_3	341×10^{-6}
H_2O	balance

ND = not detected

Table 22. SUMMARY OF RESULTS OF BY-PRODUCTS FORMATION EXPERIMENTS

Ammonia Formation

Experiment	NH_3 Content of Condenser Off-Gas (mole fraction)
H - 34	341×10^{-6}
H - 35	346×10^{-6}
C - 27	445×10^{-6}
C - 28	520×10^{-6}

Table 23. CATALYST ACTIVITY TESTS

Chemical Analyses[a]	165445	165446	165448	165449	165450
Al_2O_3, wt %	31.7	32.2	21.6	31.1	31.4
Na_2O , wt %	0.91	0.90	0.92	0.90	0.87
SO_4 , wt %	0.067	0.050	0.057	0.051	0.050
Fe , wt %	0.34	0.35	0.33	0.33	0.34
Re_2O_3, wt %	2.20	2.17	2.19	2.14	2.17
C , wt %	0.93	0.01	0.02	0.02	0.01
Ni , ppm	654	685	660	650	678
V , ppm	123	135	145	140	151
Cu , ppm	11	11	11	11	11
TV , wt %	2.24	0.85	2.79	1.63	0.85
Physical Analyses[a]					
SA , m^2/gm	125	121	130	122	124
PV , cc/gm	0.36	0.35	0.36	0.36	0.3
ABD, gm/cc	0.82	0.78	0.79	0.82	0.80
Part. Size Dist.[a]					
0-20 μ	0	0	0	0	0
0-40 μ	10	4	1	4	4
0-80 μ	94	64	64	82	65
APS, μ	68	72	73	67	72
Davison Microactivity	69.5	69.3	67.6	71.4	68.8
CPF	0.73	0.87	0.87	0.76	0.70
GPF	2.80	4.65	3.21	4.52	2.59

where TV = total volatiles wt %
SA = surface area, m^2/gm
PV = pore volume, cc/gm
ABD = apparent bulk density, gm/cc
M = microns
CPF = carbon production factor
GPF = gas production factor
APS = average particle size, microns

[a] ppm = 10^{-4} % wt
$m^2/gm = 10^3 m^2/kg$
$cc/gm = 10^{-3} m^3/kg$
$gm/cc = 10^3 kg/m^3$
$\mu = 10^{-6} m$

Samples 165448, 165449 and 165450 were both steam stripped and regenerated in our catalyst test unit. While the steam stripping conditions varied for each sample, and were 797 K (975°F), 783 K (950°F), and 761 K (910°F), respectively, 2.39×10^5 Pa at (20 psig) for 900 s (15 minutes), the regeneration of all three samples was done at the same conditions as those for sample 165446.

According to Mr. Warren Letzsch from Davison Chemical Div., W. R. Grace Co., "The samples appear to be representative commercial products that had a higher than average nickel level. This is reflected in the gas producing factor (GPF) of the microactivity test which normally runs under 2.0." Mr. Letzsch also concludes that "a comparison of the first two samples shows that our regeneration technique removes virtually all of the carbon (coke) without doing any damage to the catalyst. The chemical and physical analyses are virtual duplicates with the exceptions, of course, of the carbon and TV [total volatiles] analyses. Stripping at the relatively mild conditions shown for the last three catalysts did not cause any significant deactivation. This is not too surprising when one considers that many commercial strippers run with almost 100% steam at 783-811 K (950-1000°F) for up to several minutes. As the data indicate, no real changes occurred in either the chemical or physical analyses. There is no evidence of pore sintering, and sieve stability appears to be excellent. Our microactivity test runs plus or minus two numbers, so we would conclude that all of the samples (even the 67.6 volume percent conversion) are within experimental error of the base catalyst.*

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*Results of Davison microactivity test differing by ± 2.0 are within experimental error of the test and do not indicate change in catalyst activity.

Mr. Letzsch, who is a recognized authority on fluid catalytic cracking catalysts and their production and application also suggested that these initial tests appear to be very encouraging and that further work is fully justified. He also recommended that before a final design is set, steam tests lasting several days should be undertaken.

4.3 DISCUSSION OF RESULTS

Over 160 FCC spent catalyst steam stripping experiments were performed on a total of nine spent catalyst samples. These samples were obtained from five U.S. refineries situated in various geographical locations. As far as we could determine at the time of sample collection, the refineries operated on crude oils with sulfur contents ranging from 0.25% to 1.0% by weight. The sulfur content of the FCC feedstocks for the catalyst samples ranged from 0.5% to 1.9%. Some of the FCC feedstocks did receive pretreatment (hydro-desulfurization) prior to processing in the catcracker.

The steam stripping conditions to which these catalyst samples were exposed were outlined in Section 4.2.2.

The technical feasibility of steam stripping of spent FCC catalyst was demonstrated and it was shown that equivalent regenerator SO_x emissions of 2.0×10^{-4} mole fraction (200 vppm) SO_2 or lower are feasible. Most of the catalysts exposed to steam stripping rates of 1 to 100 kg of steam per 100 kg of catalyst showed sulfur reduction that would result in sulfur oxide emissions of 2×10^{-4} mole fraction (200 vppm) in the regenerator off-gas. For three catalysts, the experiments with steam stripping revealed sulfur removal lower than that resulting in emissions of 2.0×10^{-4} mole fraction (200 vppm). However, even in these three cases (catalysts "B"-, "C"-, and "I"- series) substantial reduction in sulfur concentration on coke (40-50%) was observed with steam stripping rates of about 50 kg per 100 kg of catalyst or lower.

Actual steam stripping rate requirements are a function of many variables including catalyst type, type of feedstock, temperature, pressure, catalyst contact time, and catalyst residence time. Mathematical correlations were developed for some of the variables and are presented in Section 4.4. Because a large number of variables affect the steam stripping process (many of which cannot be quantitatively described), only an experiment can determine whether or not the contact with steam will result in a required sulfur reduction for a specific catalyst.

Essentially no uniform trend between steam stripping and sulfur reduction was found in the case of "C"- series catalyst. The reasons for this phenomenon are unknown but Dr. E. G. Wollaston of American Oil Company suggests that this might be attributed to the metal sulfides content of the coke deposits.

Examination of experimental results reveals that the steam stripping rates to obtain an equivalent regenerator SO_2 concentration of 2.0×10^{-6} mole fraction (200 vppm) can vary greatly for different types of catalysts. This fact is not totally unexpected considering all possible catalyst types, feedstock materials, feedstock pretreatments, and process operating conditions in commerical FCC units. However, the data presented by Conn and Brackin discussed in the Phase I final report (page 52) indicate that considerable steam savings can be obtained by increasing FCC capacity from pilot plant scale to commercial application. Conn and Brackin demonstrated a substantial steam reduction of 50 to 80%.

Our experiments were performed in a semi-batch fluidized bed reactor which in no case is representative of commercial or pilot scale FCC units. In addition to the reactor's small size, its operation was not continuous which is contrary to any FCC commercial unit. After placement in the reactor the catalyst sample was exposed to steam for various lengths of time. According to some theories of laminar and turbulent conditions existing in fluidized beds* the conditions in our reactor were laminar. To change these conditions we inserted a motionless mixer in our reactor but found no significant improvement in sulfur removal efficiencies. How the motionless mixer changed the laminar conditions in the experimental reactor was not determined due to the lack of theories and empirical correlations applicable to such systems.

Nevertheless, the semi-batch operation of our reactor, significant reactor start-up and shut-down times, possible wall effects, and reactor size should be considered as important factors that would tend to decrease the efficiency of steam-catalyst contacting. Since practical observations were made in the past and showed significant improvements in sulfur removal efficiencies by going from pilot to commercial scale, the improvements of the efficiencies observed in a laboratory scale reactor are even more likely.

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*Beňa, J., J. Ilavský, E. Kossaczský, and L. Neužil. Changes of the Flow Character in a Fluidized Bed. Collect. Czech. Chem. Commun. (Prague). 28:293-308, 1963.

The effect of steam stripping upon by-product formation was also determined during this program. The data indicate that there is no appreciable formation of sulfur compounds other than the hydrogen sulfide. Some carbonyl sulfide (COS) and carbon disulfide (CS_2) were found in the stripper off-gas but the maximum concentrations amounted to less than 0.5% by volume of the total sulfur concentration in the gas stream.

Other by-products, namely hydrocarbons, were also detected in the stripper off-gas. These were present in relatively low concentrations except for methane which often exceeded the H_2S concentration on molar basis.

Heavier hydrocarbons in stripping steam condensate were detected as total organic carbon (TOC). A linear correlation on logarithmic paper was observed between steam stripping rate and TOC concentration as illustrated in Figure 17. The concentration consistently decreases with an increasing steam stripping rate. The absolute amount of hydrocarbons found in the stripper effluent condensate, however, does not seem to vary. This suggests that only limited and fixed amounts of hydrocarbons may be stripped off the catalyst with additional amounts of steam diluting the condensate stream.

This also indicates that essentially all strippable hydrocarbons will leave the catalyst with first steam in the very initial phase of catalyst steam contacting. Thus, separating this first steam that carries most of the hydrocarbons may reduce hydrocarbon concentrations in the steam condensate.

Table 24 presents some approximations of maximum concentrations for the stripper off-gas condensate based on results obtained in our experimental program. Table 25 summarizes wastewater loadings for typical refinery operations. Comparing the data in Tables 24 and 25, we can conclude that refineries are currently treating effluents which have waste loadings similar to or higher than those produced by steam stripping. Hence, no new control technology will be needed to solve expected water pollution problems. Expansion of the existing wastewater treatment facilities may be required, however, due to the increased wastewater flow. Some carbonyl sulfide may be present in the steam condensate. It appears that the amount of COS formed in steam stripping depends upon the type of catalyst and probably its history. Consequently, the COS concentration should be determined experimentally. Also, the effect of acidity of steam condensate on the amount of dissolved COS is not known. This would be important if the steam condenser is operated in the manner described in Section 5.3.

Presence of ammonia in the stripper off-gas was also observed. Its formation occurs apparently in the same manner as that of hydrogen sulfide (Phase I report, pages 58 & 59). Actually, all of the ammonia present in the stripper off-gas dissolved in the condensate.

Several tests were performed to determine the effect of spent catalyst steam stripping upon catalyst deactivation. Samples of spent catalyst which had been steam stripped and then regenerated in our laboratory were sent for analysis to Warren Letzsch of W. R. Grace Company, a manufacturer of FCC catalyst. The results of the analyses performed

Table 24. ANALYSIS OF STRIPPER OFF-GAS CONDENSATE

Stripper Operating Conditions

Temperature = 811 K (1000°F)
Pressure = 2.39×10^5 Pa (20 psig)
Steam stripping rate = 6 kg H_2O/100 kg catalyst

Condensate Composition (no sulfide separation)

Component	Concentration (kg/m^3)
Total organic carbon (TOC)	1.300[b]
Biological oxygen demand[a]	2.600
Ammonia	0.420

[a] Calculated from TOC ($BOD_5 = 2 \times TOC$).
[b] No hydrocarbon recovery was assumed.

Table 25. REFINERY WASTEWATER LOADINGS FOR TYPICAL REFINING TECHNOLOGY[a]

Fundamental Process	BOD$_5$ (kg/m^3)	Water Flow Rate gal/bbl feed	Water Flow Rate m^3/m^3	Phenols, (kg/m^3)	Sulfides, (kg/m^3)
Crude oil and product storage	0.30	0.4	0.0095	b	b
Crude desalting	1.20	0.2	0.0048	0.0060	1.20
Crude fractionation	0.0005	50	1.19	0.0024	0.0024
Thermal cracking	0.060	2	0.048	0.012	0.060
Catalytic cracking	0.040	30	0.714	0.080	0.012
Reforming	t	6	0.143	0.014	0.020
Polymerization	0.0026	140	3.33	t[c]	0.0086
Alkylation	0.0020	60	1.43	0.0003	0.020
Solvent refining	b	8	0.190	0.043	t[c]
Dewaxing	2.60	23	0.548	0.008	t[c]
Hydrotreating	0.240	1	0.024	t[c]	0.240
Deasphalting	b	b	b	b	b
Drying and sweetening	0.150	40	0.952	0.030	b
Wax finishing	b	b	b	b	b
Grease manufacturing	b	b	b	b	b
Lube or finishing	b	b	b	b	b
Blending and packaging	b	b	b	b	b

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[a] Jones, H. R., Pollution Control in the Petroleum Industry, Noyes Data Corporation, Park Ridge, N. J., 1973.

[b] Data not available for reasonable estimate.

t[c] Trace

indicated that the exposure of catalyst to steam for 900 s (15 minutes) at 2.39×10^5 Pa (20 psig) and temperatures ranging from 755 to 797 K (900 to 975°F) did not cause any catalyst deactivation.

In summation, the experimental work did not reveal any information that would require changing conclusions drawn in the Phase I final report (pages 39-43). The steam stripping of spent catalyst is a technically feasible method of reducing FCC regenerator SO_x emissions. This method does not form large quantities of undesirable compounds but forms H_2S which can be converted to saleable grade sulfur at existing refineries. Also, it appears that steam stripping will cause no drastic catalyst deactivation as suspected at several petroleum refineries.

4.4 DATA REGRESSION ANALYSIS

The experimental data presented in the previous section were analyzed using regression analysis techniques. The empirical correlation form which best fits the data obtained from all catalysts is presented below. For some catalysts, experimental data may seem to fit other correlations better than that in Equation (3). Our intent, however, was to obtain a general correlation that would represent best the reaction phenomena for all catalysts.

$$\ln(SO_2)_{out} = \ln(SO_2)_{in} - k\ T_s(SSR) \qquad (3)$$

where $(SO_2)_{out}$ = equivalent SO_2 concentration (vppm) after catalyst steam stripping (equivalent to residual sulfur content on coke)*,+

$(SO_2)_{in}$ = equivalent regeneration SO_2 concentration (vppm) before catalyst steam stripping (equivalent to initial sulfur content on coke)*,+

k = proportionality constant

T_s = steam residence in catalyst bed in stripper (minutes)++

(SSR) = steam stripping rate (kg H_2O/100 kg of catalyst)

If we define fractional sulfur removal efficiency as

$$X = \frac{(SO_2)_{in} - (SO_2)_{out}}{(SO_2)_{in}} \tag{4}$$

or

$$X = 1 - \frac{(SO_2)_{out}}{(SO_2)_{in}} \tag{5}$$

Equation (3) becomes

$$\ln(1-X) = -k\,T_s(SSR) \tag{6}$$

---

*Both concentrations were calculated according to the procedure outlined in Section 4.2.

+vppm = 10^6 mole fraction

++1 min = 60 seconds

The data used in regression analysis for each of the catalysts and their manipulation to determine the constants of Equation (3) are summarized in Table 26. This table also includes additional calculated values according to the following nomenclature:

Equation (3) can be simply transcribed in the form

$$Y = A + BX$$

where $Y = \ln(SO_2)_{out}$

$X = T_s(SSR)$

$A = \ln(SO_2)_{in}$

$B = -k$

and $\overline{X}$ = mean of X

σ_X = standard deviation of X

TT = T - test

EE = standard error of estimate

R = simple correlation coefficient

At the end of this section a summary of correlation equations obtained for each catalyst from regression analysis is presented and their agreement with experimental results demonstrated (Figures 18 through 25).

Table 26. STEAM STRIPPING DATA REGRESSION ANALYSIS

Experiment	T_s	(SSR)	(SO_2) out	X	Y	$\overline{X}$	σ_x	TT	R^2	A	B
B 3	1.23	5.31	979	6.5313	6.8865	28.904	18.210	-2.105	0.5257	6.86	5.26×10^{-3}
5	0.791	32.8	732	25.945	6.5958						
7	1.36	17.6	951	23.936	6.8575						
8	1.48	39.8	721	58.904	6.5806						
9	2.05	9.01	805	18.471	6.6908						
11	1.81	21.9	769	39.639	6.6451						
C 6	0.888	7.31	592	6.4913	6.3835	120.26	160.28	-3.489	0.6791	6.33	5.26×10^{-4}
7	0.924	14.1	549	13.028	6.3081						
8	0.798	32.6	540	26.015	6.2916						
10	1.80	16.0	493	28.800	6.2005						
11	1.27	5.98	610	7.5946	6.4135						
15	1.75	7.59	559	13.283	6.3262						
17	1.48	219	460	324.12	6.1312						
18	2.43	109	483	264.87	6.1800						
21	0.631	631	471	398.16	6.1549						
D 2	1.63	17.7	385	28.851	5.9532	145.16	131.07	-2.446	0.8568	5.92	3.00×10^{-3}
5	1.32	90.5	217	119.46	5.3799						
7	1.73	166	168	287.18	5.1240						
E 6	0.880	97.8	162	86.064	5.0876	17.281	26.355	-23.067	0.9870	5.96	1.0×10^{-2}
15	1.63	6.32	362	10.302	5.8916						
16	1.48	13.3	305	19.684	5.7203						
17	1.84	5.20	361	9.5680	5.8889						
18	1.06	2.85	362	3.0210	5.8916						
21	0.937	11.6	339	10.869	5.8260						
22	0.642	1.34	393	0.8603	5.9738						
23	0.879	6.16	368	5.4146	5.9081						
24	0.826	11.8	366	9.7468	5.9026						
F 9	0.782	5.54	438	4.3323	6.0822	13.081	7.732	-4.388	0.6582	6.19	2.0×10^{-2}
10	0.927	10.5	408	9.7335	6.0113						
11	0.748	2.17	477	1.6232	6.1675						
12	1.37	6.99	416	9.5763	6.0307						
13	1.29	7.41	440	9.5589	6.0868						
14	1.72	11.1	342	19.092	5.8348						
15	1.50	9.57	377	14.355	5.9323						
17	1.88	9.87	417	18.556	6.0331						
18	1.36	9.07	290	12.335	5.6699						
22	2.63	8.20	316	21.566	5.7557						
25	2.41	11.9	256	28.679	5.5452						
27	3.29	2.30	422	7.5670	6.0450						

Table 26 continued. STEAM STRIPPING DATA REGRESSION ANALYSIS

Experiment		T_s	(SSR)	(SO_2) out	X	Y	$\overline{X}$	σ_x	TT	R^2	A	B
H	5	0.539	4.01	175	2.1614	5.1648	72.496	106.23	-4.388	0.6357	4.91	2.02×10^{-3}
	6	0.847	7.68	143	6.5050	4.9628						
	7	1.01	12.9	136	13.029	4.9127						
	8	0.81	32.1	139	26.001	4.9345						
	11	0.755	115	73	86.825	4.2905						
	12	1.10	4.37	110	4.8070	4.7005						
	13	1.78	5.38	136	9.5764	4.9127						
	14	1.60	8.95	127	14.320	4.8442						
	15	1.68	17.1	144	28.728	4.9608						
	16	1.82	31.6	119	57.512	4.7791						
	17	1.84	65.0	95	119.60	4.5539						
	18	1.55	185	84	286.75	4.4308						
	19	1.42	4.67	164	6.6314	5.0999						
	21	2.42	164	63	396.88	4.1431						
	22	1.92	6.89	148	13.229	4.9972						
	23	2.03	9.78	126	19.853	4.8363						
	24	2.54	15.6	106	39.624	4.6634						
	25	1.90	41.8	99	79.420	4.5951						
	26	2.53	65.6	109	165.97	4.6914						
H[a]	28	0.803	72.0	103	57.816	4.6347	44.482	57.134	-5.447	0.4331	4.87	3.24×10^{-3}
	29	0.777	179	85	139.08	4.4427						
	30	0.765	18.2	111	13.923	4.7095						
	31	0.663	10.5	102	6.9615	4.6250						
	32	0.525	8.81	180	4.6253	5.1930						
I	5	0.517	4.19	477	2.1662	6.1675	74.335	90.186	-4.362	0.7040	6.03	1.82×10^{-3}
	8	0.596	29.1	386	17.344	5.9558						
	9	0.648	53.4	361	34.603	5.8889						
	10	0.669	194	304	129.79	5.7170						
	11	0.648	13.4	422	8.6832	6.0450						
	12	2.40	3.99	460	9.5760	6.1312						
	13	1.21	31.6	400	38.236	5.9915						
	14	1.48	97.1	304	143.71	5.7170						
	15	1.58	182	274	287.56	5.6131						
	16	1.46	49.1	300	71.686	5.7038						

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[a] This experiment was performed at 900°F; all others were performed at 1000°F.

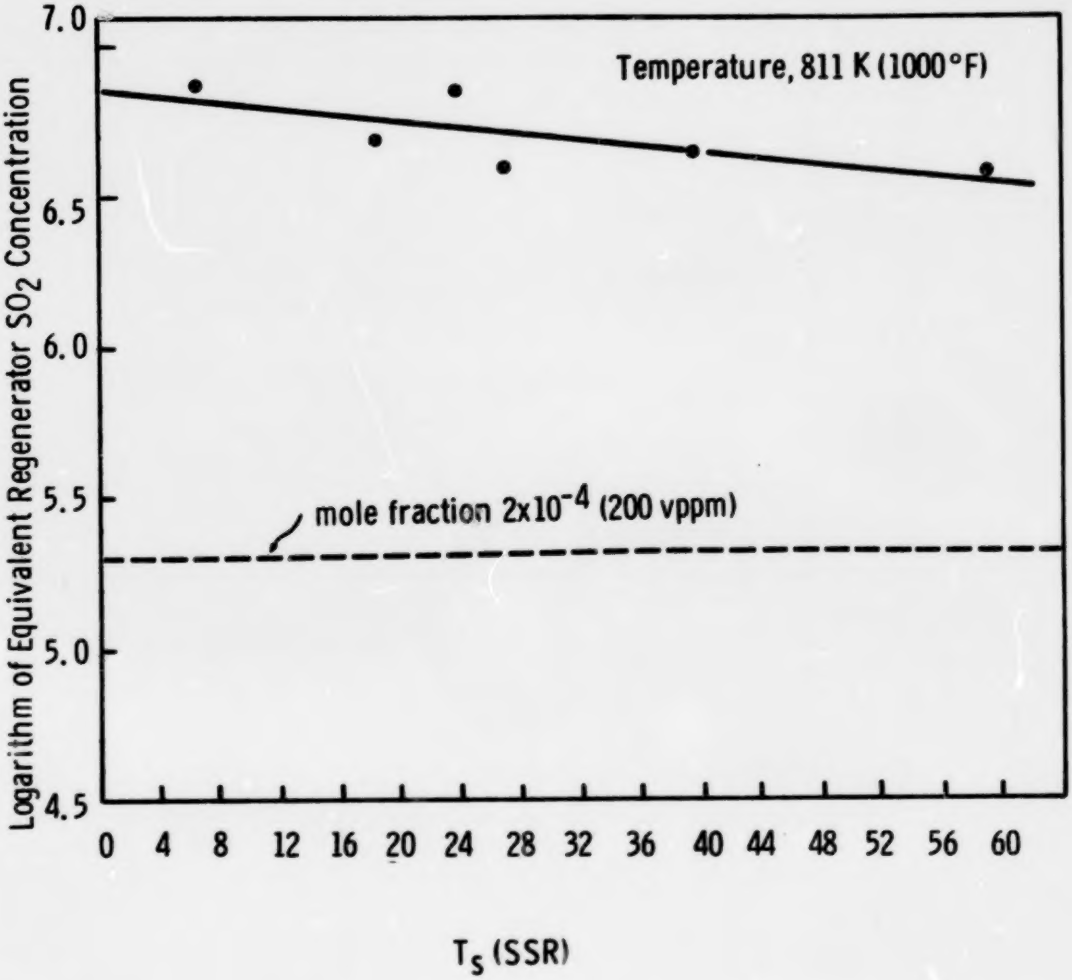

Figure 18. Summary of results of the regression analysis performed on B-Series catalyst

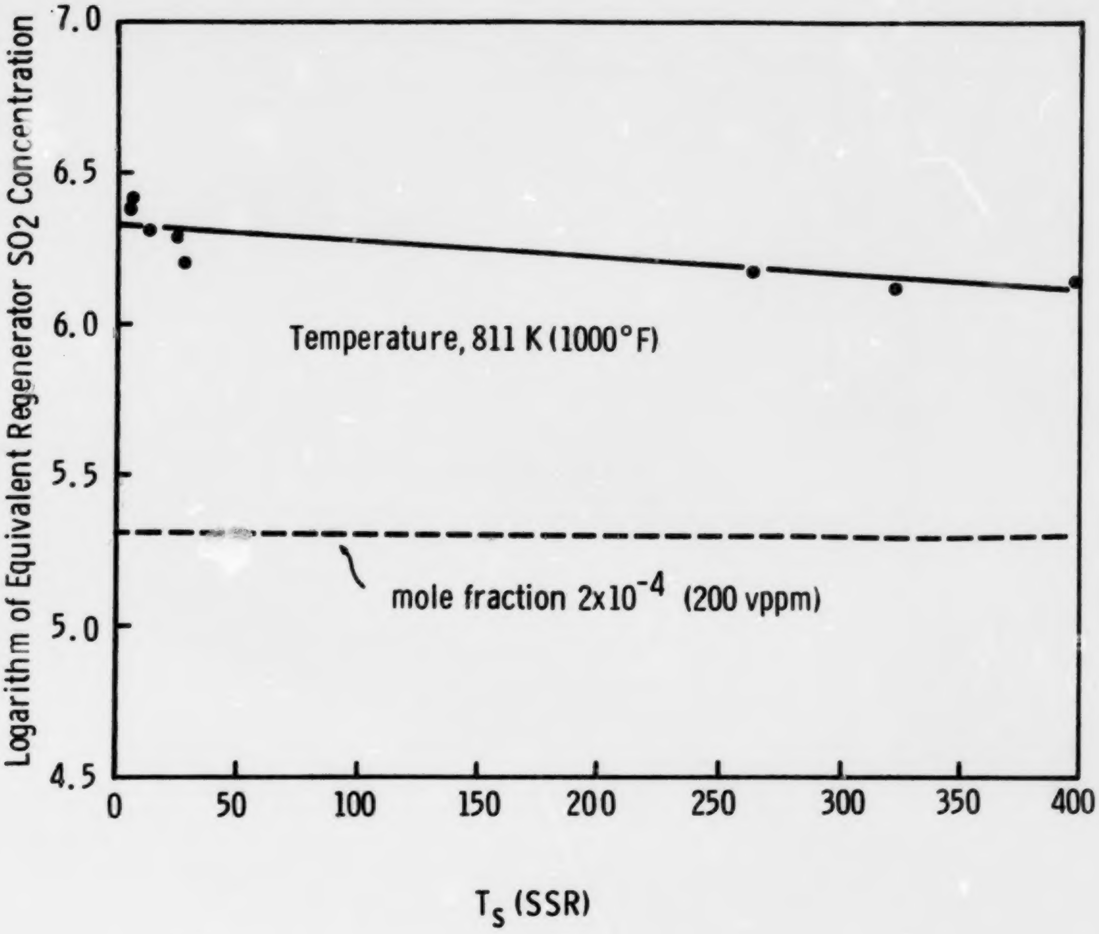

Figure 19. Summary of results of the regression analysis performed on C-Series catalyst

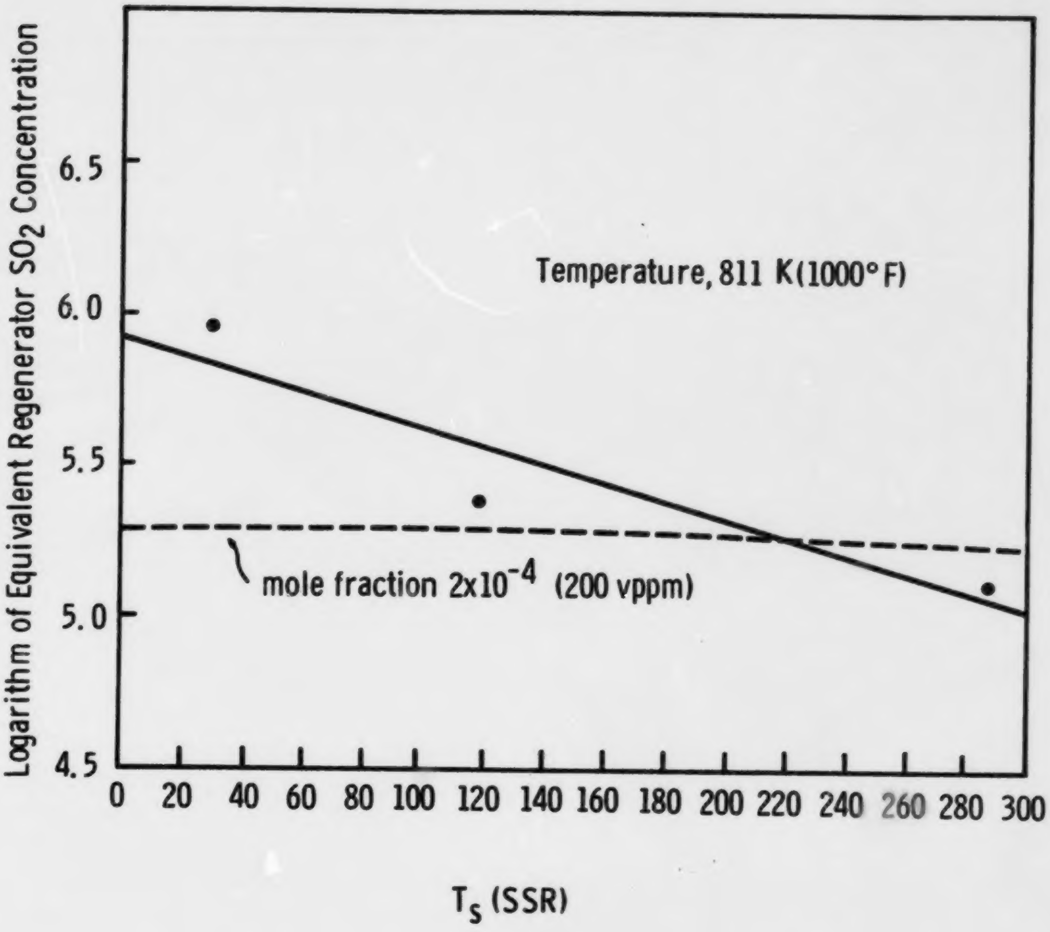

Figure 20. Summary of results of the regression analysis performed on D-Series catalyst

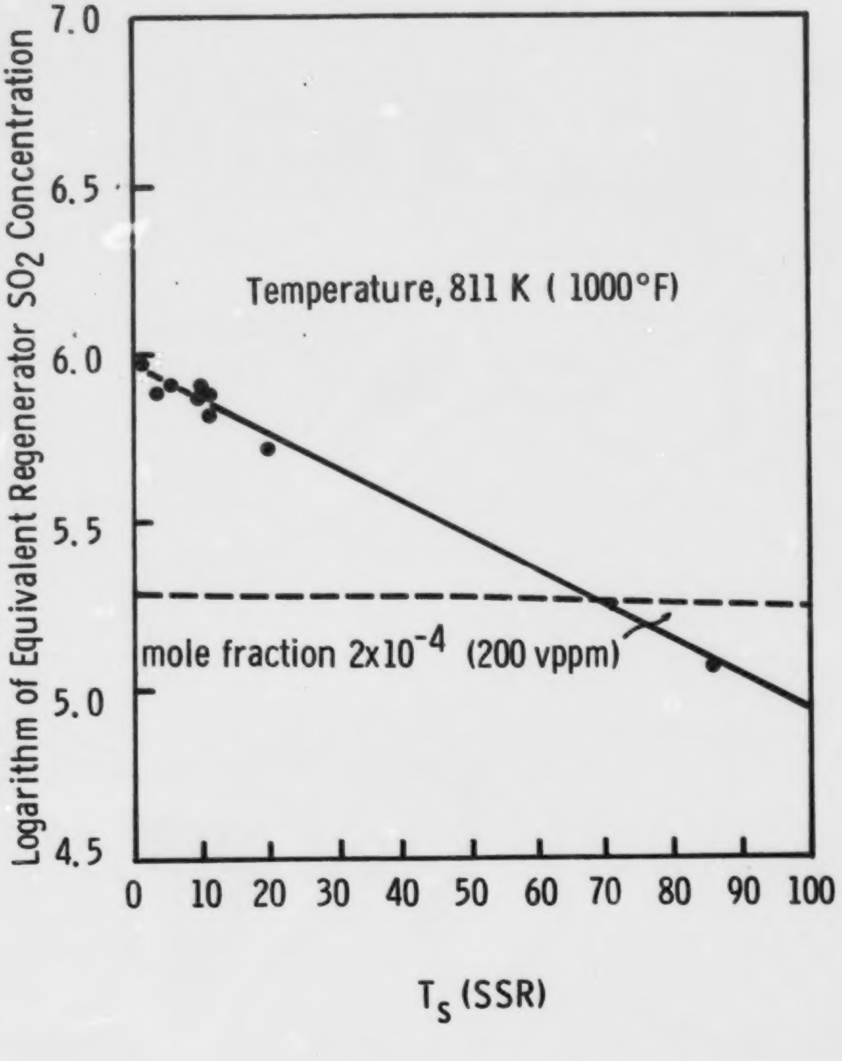

Figure 21. Summary of results of the regression analysis performed on E-Series catalyst

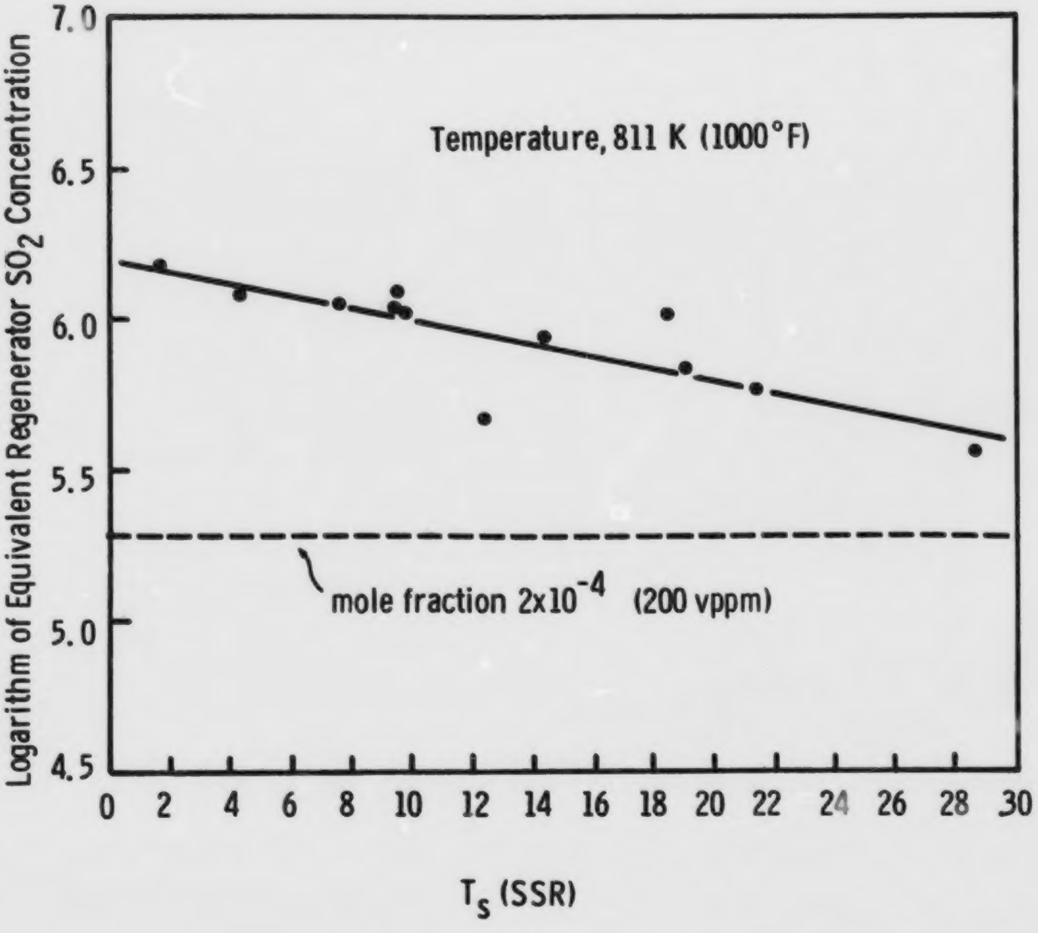

Figure 22. Summary of results of the regression analysis performed on F-Series catalyst

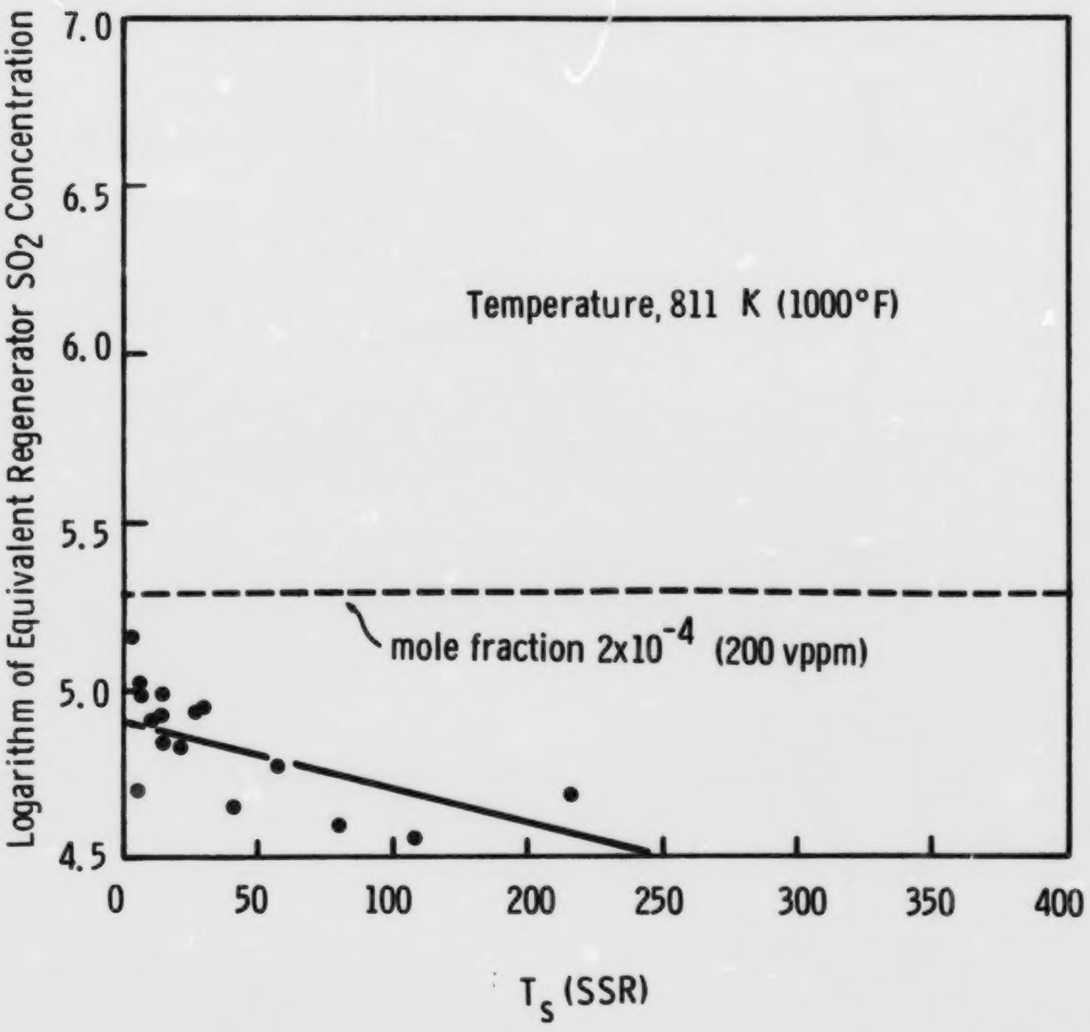

Figure 23. Summary of results of the regression analysis performed on H-Series catalyst

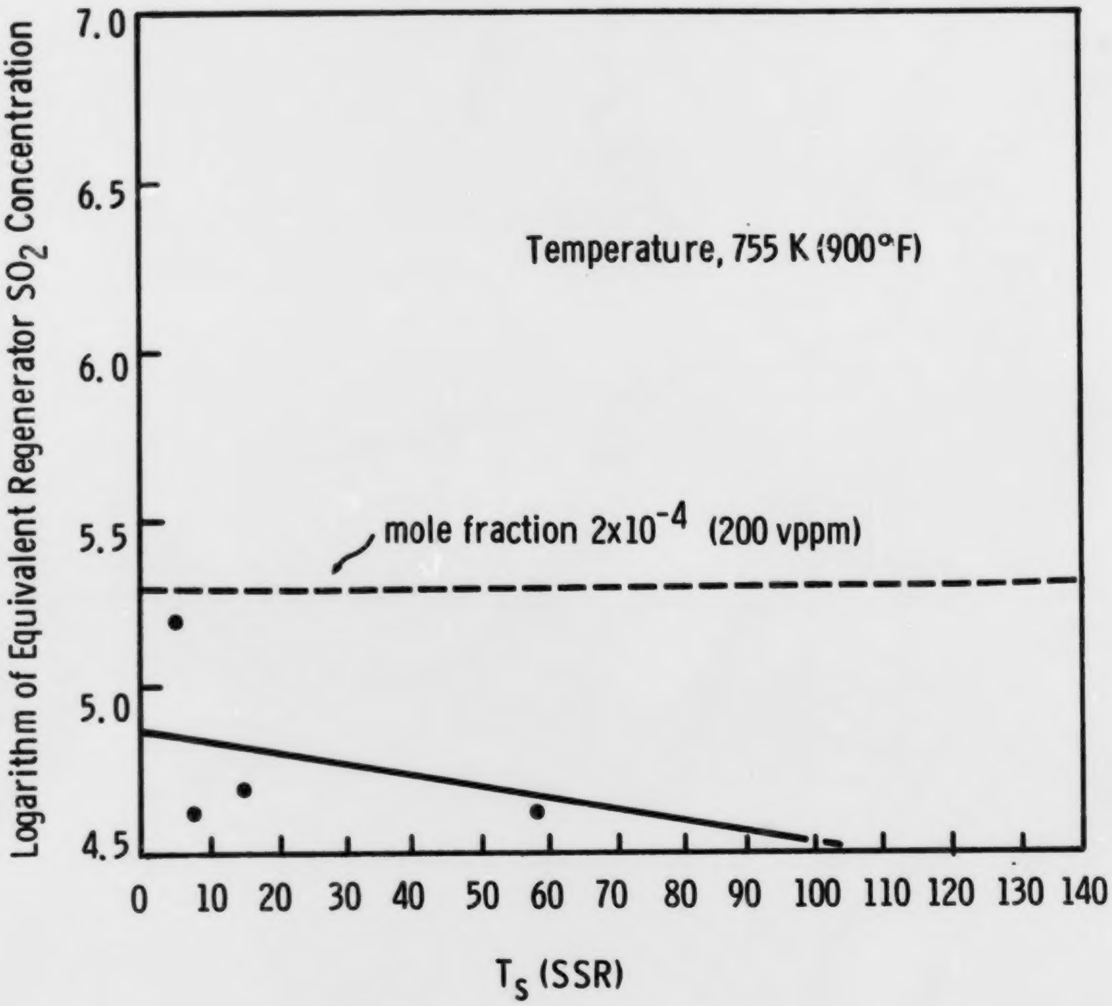

Figure 24. Summary of results of the regression analysis performed on H-Series catalyst

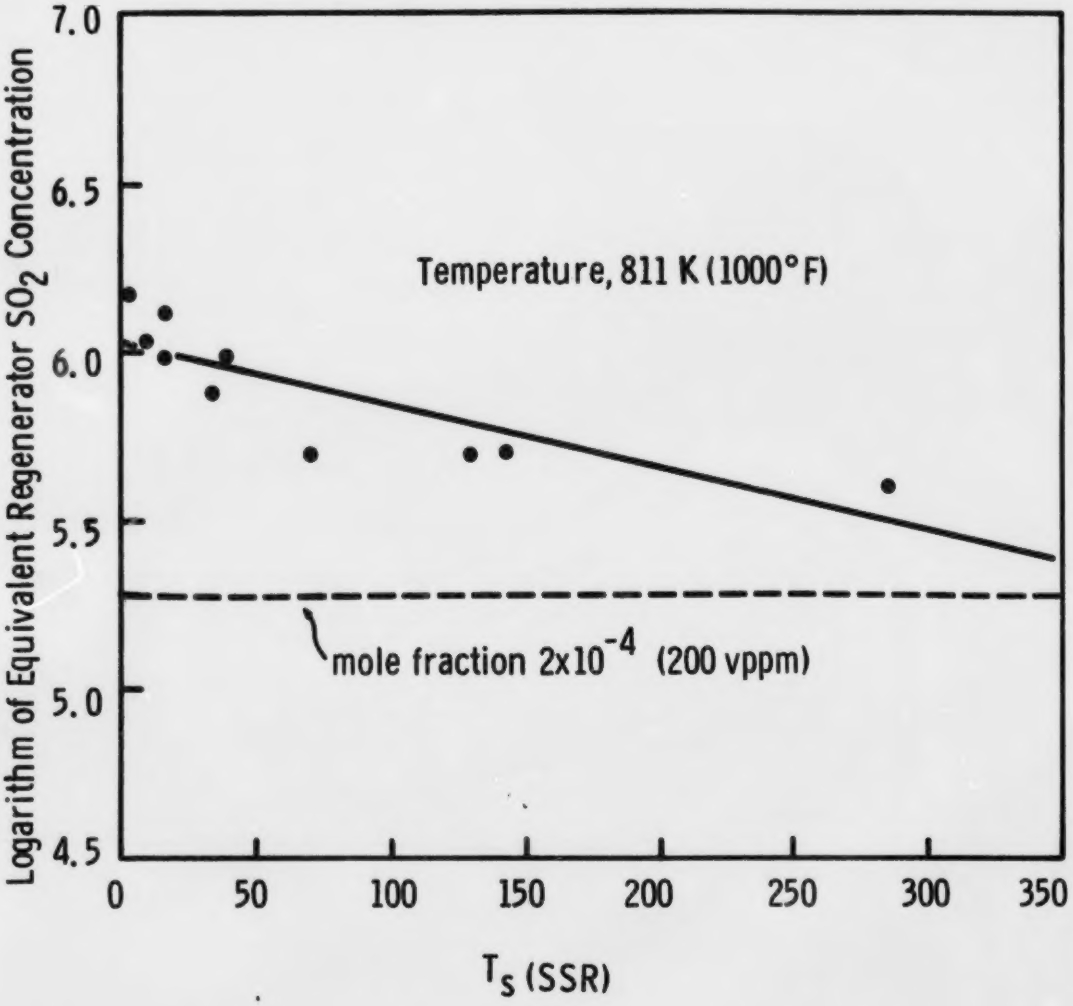

Figure 25. Summary of results of the regression analysis performed on I-Series catalyst

Summary of Correlation Equations

Catalyst	
B	$\ln(SO_2)_{out} = 6.86 - 5.26\text{x}10^{-3}\ T_s\ (SSR) \pm 3.04\%$
C	$\ln(SO_2)_{out} = 6.33 - 5.262\text{x}10^{-4}\ T_s\ (SSR) \pm 1.98\%$
D	$\ln(SO_2)_{out} = 5.92 - 3.00\text{x}10^{-3}\ T_s\ (SSR) \pm 8.28\%$
E	$\ln(SO_2)_{out} = 5.96 - 1.0\ \text{x}10^{-2}\ T_s\ (SSR) \pm 0.114\%$
F	$\ln(SO_2)_{out} = 6.19 - 2.0\ \text{x}10^{-2}\ T_s\ (SSR) \pm 3.94\%$
H^a	$\ln(SO_2)_{out} = 4.91 - 2.02\text{x}10^{-3}\ T_s\ (SSR) \pm 7.04\%$
H^b	$\ln(SO_2)_{out} = 4.87 - 3.24\text{x}10^{-3}\ T_s\ (SSR) \pm 10.4\%$
I	$\ln(SO_2)_{out} = 6.03 - 1.82\text{x}10^{-3}\ T_s\ (SSR) \pm 3.84\%$

a = 1000°F
b = 900°F

The percent error for each equation was calculated at the mean $\overline{X}$ for interval of 2 times EE, which should include 95% of all data in regression analysis. Table 27 summarizes the term T_s (SSR) calculated from the correlation equations to obtain 200 vppm equivalent regenerator SO_2 concentrations.

Table 27. STEAM STRIPPING REQUIREMENT FOR SULFUR REDUCTION TO 200 vppm

Catalyst	T_s (SSR)
B	297
C	1961
D	207
E	66.2
F	45.6
H	136 vppm initially
I	402

5. STEAM STRIPPING PROCESS DESIGN

Applying the steam stripping process for refinery FCC unit regenerator SO_x control requires several processing steps. In the Phase I final report several process alternatives were proposed and one of these (Option 1) was evaluated in detail. The laboratory development program (Phase II) did not reveal any evidence that would require a modification of the Option 1 alternative. However, in many cases new technical information was obtained or generated which enables a better understanding of the individual process steps for optimization of equipment design. In this section, discussions on individual processing steps and technical background information are presented and applied to proper processing equipment design.

5.1 CATALYST STEAM STRIPPER DESIGN

In applying the Option 1 process alternative in the Phase I final report, we indicated that several types of equipment may be used to contact the spent catalyst with steam. Namely, these are:

- Fluidized bed catalyst stripper (similar to existing FCC unit reactors and regenerators)
- Counter-current, stagewise contacting (similar strippers presently used in refineries and illustrated in Figure 3, Phase I final report, p. 17)

- Co-current, plug-flow contacting (similar to riser reactor concept applied to FCC hydrocarbon cracking; this concept would require a catalyst disengagement step following the stripper)

These concepts are discussed below.

Computer analysis of the experimental data obtained in the laboratory development program revealed that the removal efficiency of sulfur from the coke on spent catalyst is a function of several variables. Specifically, the mathematical correlations containing the steam stripping rate and time during which the catalyst is exposed to steam were developed for all catalysts tested. The correlations were presented in Section 4.2.3 and appeared to have the general form of Equation (6) for all catalysts.

The constant in the Equation (6) includes factors such as temperature, pressure, type of catalyst, type and concentration of sulfur compounds on coke, and type of equipment in which the catalyst steam contacting takes place. The equation may become very useful in further development of the steam stripping process. Its further extrapolation to large scale units, however, will have to be verified.

5.1.1 Semi-Batch Fluidized Bed Reactor

The following theoretical discussions will further clarify the meaning and interpretation of the constant in Equation (6). As a starting point, we will assume that the catalyst steam stripper can schematically be depicted in the manner shown in Figure 26. The operation of this stripper is similar to our experimental reactor.

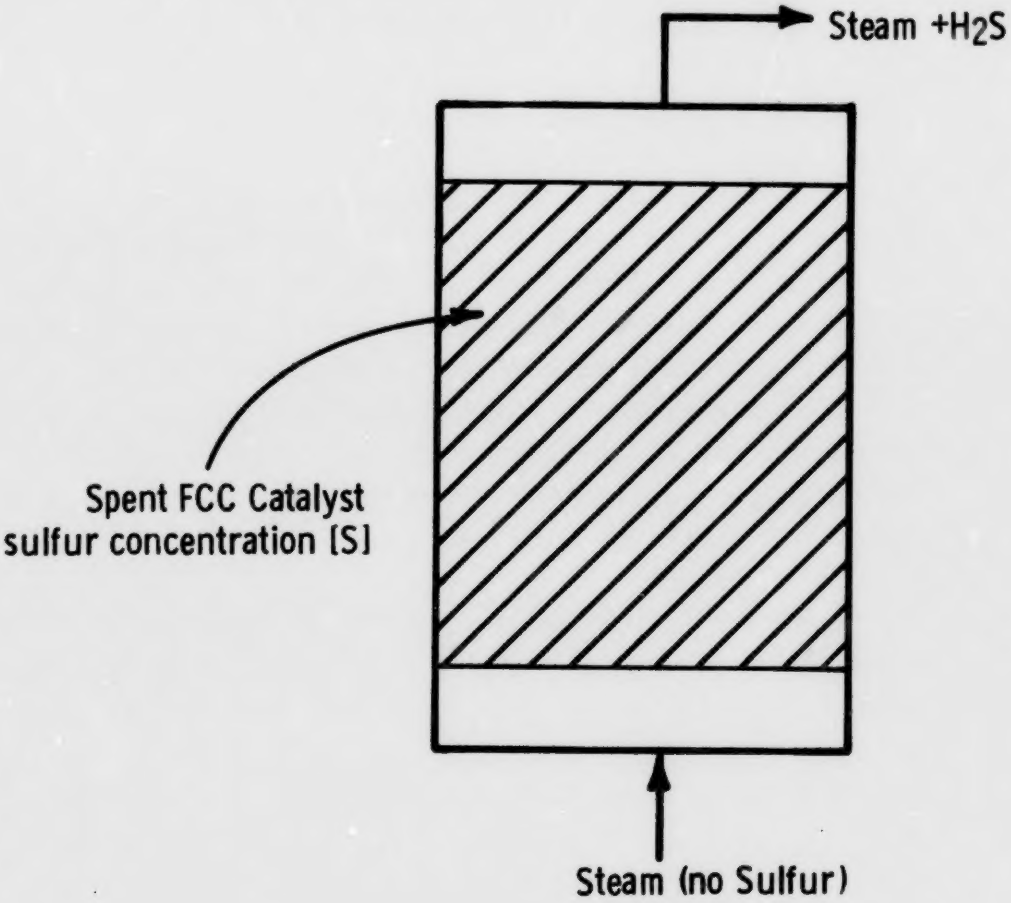
Steam +H2S
Spent FCC Catalyst
sulfur concentration [S]
Steam (no Sulfur)

Figure 26. Schematic of spent catalyst steam stripper

Assuming that the rate of sulfur removal from spent FCC catalyst is a function of the sulfur content of coke and amount of steam to which the catalyst is exposed, the following correlation may be used to describe this relationship.

$$-\frac{dS}{dt} = k\,[S]^a\,[H_2O]^b \qquad (7)$$

where $[S]$ = concentration of removable sulfur on spent catalyst

$[H_2O]$ = concentration of steam

a,b = constants expressing the order of stripping reactions

k = proportionality constant in any of the equations below, this constant must be experimentally determined

t = time

Previous work related to catalyst coke composition revealed that the sulfur on the coke can be in various forms. This was also discussed in the Phase I final report, Section 5.2.2. Mathematically we can describe the sulfur forms on the catalyst as follows:

$$S_T = S_R + S \qquad (8)$$

where S_T = total sulfur on spent catalyst

S_R = residual sulfur

S = sulfur removed by steam stripping

Thus, in Equation (7) above, [S] represents the sulfur that can be removed by contacting the catalyst with steam.

In order to integrate Equation (7), certain boundary conditions must be defined. These are:

$$\begin{aligned} [S] &= S_1 \quad \text{at} \quad t = 0 \\ [S] &= S_2 \quad \text{at} \quad t = T_c \\ [S] &= S_R \quad \text{at} \quad t = \infty \end{aligned} \qquad (9)$$

where T_c = catalyst residence time in the stripper, or time during which the catalyst is exposed to steam

In operation with excess steam we can assume $[H_2O] \neq f$ (time, conversion). Integrating Equation (7) and applying the above boundary conditions we obtain

$$\frac{1}{1-a}\left[(S_2)^{(1-a)} - (S_1)^{(1-a)}\right] = -k\left[H_2O\right]^b T_c \qquad (10)$$

for $a \neq 1$, and

$$\ln [S_2] - \ln [S_1] = -k\,[H_2O]^b\,T_c \qquad (11)$$

for $a = 1$.

In Equations (10) and (11), S_1 represents the initial sulfur concentration on coke and S_2 the sulfur concentration on coke after the time T_c.

To better compare the experimental results with these correlations we will define a new variable, X, representing the fraction of sulfur removed from the catalyst in time T_c.

$$X = \frac{[S_1] - [S_2]}{[S_1]} \qquad (12)$$

This variable is identical to the one defined by Equation (4), Section 4.4.

Substituting Equation (12) into Equations (10) and (11) we obtain

$$\frac{(S_1)^{(1-a)}}{1-a}\left[(1-X)^{(1-a)}-1\right] = -k\ [H_2O]^b\ T_c \qquad (13)$$

and

$$\ln\ (1-X) = -k\ [H_2O]^b\ T_c \qquad (14)$$

Further modification will be made with the right side of Equations (13) and (14). First, we will multiply the terms on the right side by T_s/T_s, where T_s is the steam residence time in catalyst stripper.

$$-k\ [H_2O]^b\ T_c = -k\ [H_2O]^b\ T_c\ \frac{T_s}{T_s} \qquad (15)$$

Examination of the Equation (15) will reveal that new terms which were actually measured in our experimental studies can be introduced in Equations (13) and (14).

The steam residence time T_s can also be expressed as a function of reactor volume V_R:

$$T_s = \frac{V_R}{V_s} = \frac{V_R\ \rho_s}{\overline{M}_s} \qquad (16)$$

where V_s = volume rate of steam (m^3/s)
V_R = reactor volume (m^3)
ρ_s = steam density (kg/m^3); in excess conditions equal to steam concentration
$\overline{M}_s$ = steam mass rate (kg/s)

Similarly, we can express catalyst mass in the reaction as

$$C = V_R \rho_c \tag{17}$$

where C = mass of catalyst used for an experiment (kg)
ρ_c = catalyst density in the reactor (kg/m^3)

Substituting T_s and using Equation (17) we will introduce a steam-to-catalyst ratio term in Equation (15).

$$-k\,[H_2O]^b\,T_c = -k[H_2O]^b\,\frac{T_c\,T_s\,\rho_c}{\rho_s} \times \frac{\overline{M}_s}{C} \tag{18}$$

The $\overline{M}_s/C$ ratio can also be expressed in terms of steam stripping rate (SSR)

$$\frac{\overline{M}_s}{C} = \frac{(SSR)}{100XT_c} \tag{19}$$

which after modification of Equation (18) will produce

$$-k\,[H_2O]^b\,T_c = -k\,[H_2O]^b\,T_s\,(SSR)\,\frac{\rho_c}{\rho_s} \tag{20}$$

Equation (20) is only the right side of the original Equations (13) and (14). Let us combine these Equations with Equation (20).

$$\frac{(S_1)^{(1-a)}}{1-a}\left[(1-X)^{(1-a)}-1\right] = -k\,[H_2O]^b\,\frac{\rho_c}{\rho_s}\,T_s\,(SSR) \qquad (21)$$

for $a \neq 1$, and

$$\ln(1-X) = -k\,[H_2O]^b\,\frac{\rho_c}{\rho_s}\,T_s\,(SSR) \qquad (22)$$

for $a = 1$.

Equations (21) and (22) are in a general form which can be used to determine the effect of steam stripping rate and steam residence time on sulfur removal efficiency. The right side can be further simplified according to the following assumptions. If the effect of steam on the sulfur removal is of the first order, the constant $b = 1$ (our regression analysis of experimental data showed that this is a reasonable assumption as presented in Section 4.4) and the steam concentration will cancel out with actual steam density. This is possible because of high excess of steam. Also, if the catalyst reaction density ρ_c is considered constant over a narrow range of operating conditions and its change is expressed in terms of steam residence time T_s (normally $T_s \times \rho_c$ = constant), the term ρ_c may be included into the proportionality constant k. Thus, our final simplified equations will become

$$\frac{(S_1)^{(1-a)}}{1-a}\left[(1-X)^{(1-a)}-1\right] = -k\,T_s\,(SSR) \qquad (23)$$

for $a \neq 1$, and

$$\ln(1-X) = -k\,T_s\,(SSR) \qquad (24)$$

for a = 1, or

$$X = 1-e^{-k\ T_S\ (SSR)} \qquad (25)$$

The regression analysis of experimental data produced a correlation that very well satisfies Equation (24), which suggests that the effect of sulfur on the sulfur removal rate is also of the first order.

Together, Equations (23) and (24) can be used to predict the sulfur removal efficiency as a function of steam stripping rate and steam residence time. Assuming that the proportionality constant k will represent the effects of temperature, pressure, type of catalyst, and type of steam-catalyst contacting device, the applicability of the equations is further expanded.

5.1.2 Rate Controlling Factors

Several groups of data obtained for the same catalyst at equal catalyst residence times but various stripper steam velocities confirmed that the same sulfur removal can be obtained with lesser amounts of steam at lower steam velocities as long as the catalyst residence time remains the same. Several steps might be involved in bringing the sulfur on the catalyst to the form in which it is removed. Some of the steps are suggested below even though it is not known which of these steps are the controlling ones.

(1) Rate of diffusion of steam through the pores in each catalyst particle.

(2) Rate of reaction of steam and sulfur compounds on catalyst surface.

(3) Rate of diffusion of product H_2S through the pores in catalyst.

(4) Rate of diffusion of product H_2S through laminar boundary layer surrounding each catalyst particle.

Should the first or third factor be controlling, our experimental data would have shown essentially no effect of pressure on sulfur removal efficiency. Increased pressures, however, resulted in higher removal efficiency (see Figure 13).

Should the second factor be controlling, a significant difference would be expected between the data measured at different temperatures. This effect, however, can be partially compensated for or enhanced by the other effects, the fourth one in particular. Nevertheless, essentially very minimal temperature effect was observed.

A single effect of the fourth factor should indicate a significant improvement in sulfur removal with an increased turbulence in the reactor. Insertion of static mixer in the reactor should enhance the turbulence of the fluidized bed but it did not result in a remarkable improvement and consequently does not support the significance of laminar layer diffusion. However, the theory of the laminar and turbulent conditions in fluidized beds is not well defined and, as mentioned earlier, the effect of static mixer on turbulent conditions in the fluidized bed is not easy to quantify. Therefore, it is difficult to objectively evaluate the improvement of turbulent conditions in the fluidized reactor by the use of a static mixer.

In conclusion, the minimal effect of temperature and rather significant effect of pressure on sulfur removal efficiency seem to indicate that kinetics (Factor 2) is the controlling step, with elevated pressures resulting in easier sulfur removal. As a result, potential improvement in sulfur removal kinetics may be sought through an investigation of catalytic effects of trace elements normally contained in petroleum feedstocks and deposited on FCC catalyst during the cracking process. These effects were not evaluated in this program.

Better sulfur removal efficiencies were observed by going from pilot to commercial scale stripper (Phase I final report, Section 5.2.1). This seems to suggust that Factor 4 has some significance. This can be partially explained by the more uniform conditions and the minimization of wall and start-up effects in commercial units. Considering this observation, our experiments performed in a fluidized bed reactor in a rather semi-batchwise manner should not be viewed as representative of FCC commercial units since substantial wall, start-up, and mixing effects were probably present. Consequently, we feel that the experimental results observed on the laboratory scale can be significantly improved in operations of commercial size.

5.1.3 Other Stripper Designs

The petroleum industry has spent roughly 40 years developing various catalytic cracking processes. Research performed in this area has resulted in improvements in FCC reactor design, spent catalyst steam stripping for hydrocarbon removal, and spent catalyst regeneration. The experience gained in these extensive R&D efforts can be used to aid in designing spent catalyst steam strippers for the purpose of sulfur removal. Our semi-batch experimental reactor was described previously. Three other design alternatives are currently envisioned, including continuous fluidized bed, counter-current, and co-current reactors. The theoretical math model developed previously for semi-batch bed is extended below to cover the other two stripper design alternatives.

5.1.3.1 Continuous Fluidized Bed Reactor -

The same type of kinetic model development used to determine the behavior of a batch fluidized bed catalyst stripper can be used for a continuous fluidized bed catalyst stripper if we assume that the stripper behaves as a constant stirred tank reactor (CSTR), Figure 27.

Basically, we can assume that $[S] = S_2 =$ constant and no concentration change occurs in the reactor, or $\frac{dS}{dt} = 0$. The sulfur concentration entering the reactor must be equal to the concentration leaving the reactor plus the amount of sulfur reacted, or

$$S_1 - S_2 - k\, S_2{}^a\, [H_2O]^b\, T_c = 0 \qquad (26)$$

Further modification of the equation is possible:

$$\frac{S_1 - S_2}{S_2{}^a} = k\, [H_2O]^b\, T_c \qquad (27)$$

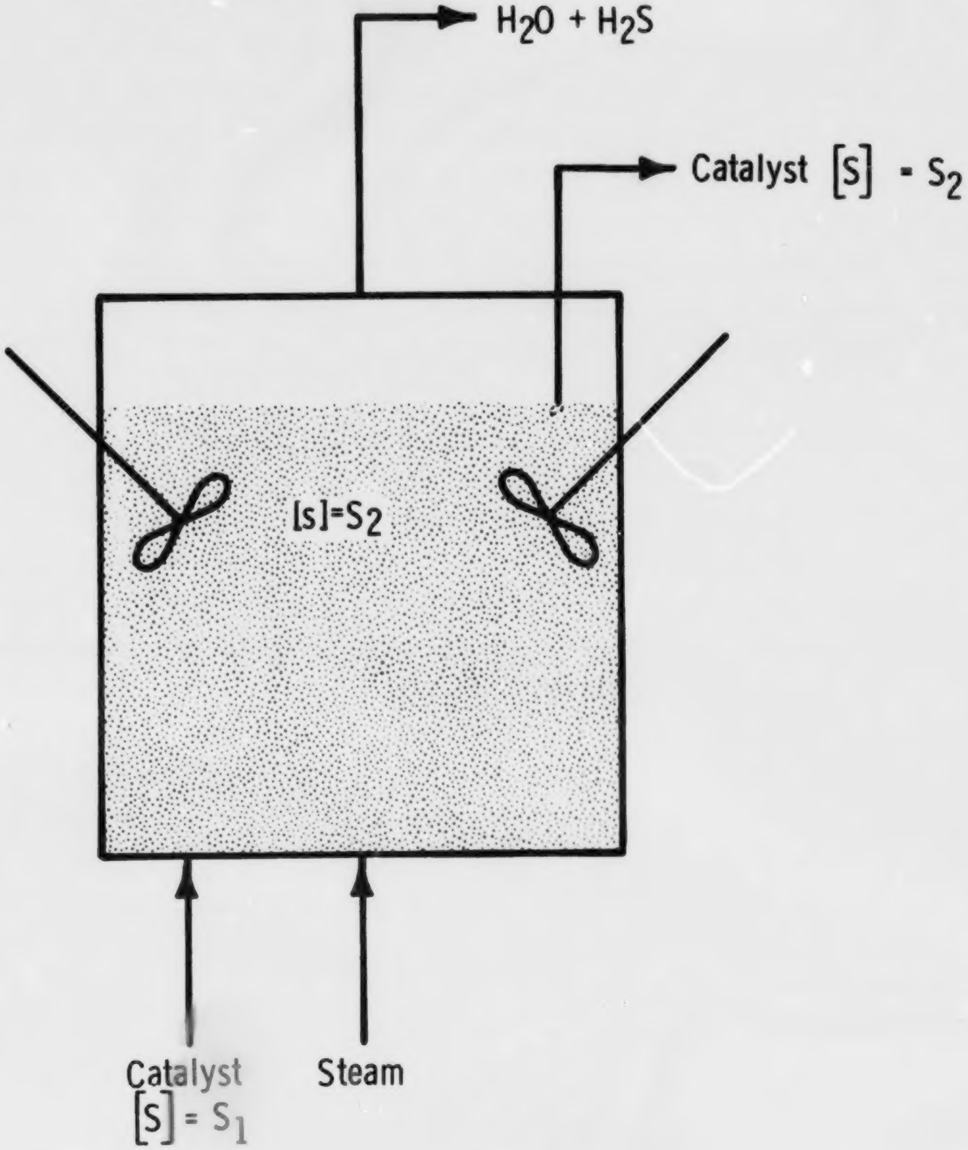
$H_2O + H_2S$
Catalyst $[S] = S_2$
$[S] = S_2$
Catalyst
$[S] = S_1$
Steam

Figure 27. Schematic of continuous fluidized bed stripper

for a + 1, and

$$\frac{S_2}{S_1} = \frac{1}{1 + k\,[H_2O]\,T_c} \tag{28}$$

for a = 1.

Considering Equation (12) and assuming a = 1, and b = 1, the sulfur removal can also be expressed as

$$1 - X = \frac{1}{1 + k\,[H_2O]\,T_c} \tag{29}$$

or

$$X = \frac{k\,[H_2O]\,T_c}{1 + k\,[H_2O]\,T_c} \tag{30}$$

Expressing T_c in the following form,

$$T_c = \frac{V_R\,\rho_c}{C} \tag{31}$$

using Equation (16) and the following Equation (32)

$$\overline{M}_{S/C} = \frac{(SSR)}{100} \tag{32}$$

we can include steam stripping rate into Equation (30)

$$X = \frac{k[H_2O]\,\frac{\rho_c}{\rho_s}\,T_s\ (SSR)}{1 + k[H_2O]\,\frac{\rho_c}{\rho_s}\,T_s\ (SSR)} \tag{33}$$

Applying the assumptions listed on page 109, we obtain the equation in its final form

$$X = \frac{k\ T_s\ (SSR)}{1 + k\ T_s\ (SSR)} \tag{34}$$

5.1.3.2 Plug-Flow Stripper -

A spent catalyst steam stripper can be operated such that the steam and catalyst pass through the stripper in the same direction, or co-currently. Co-current steam stripping is thus performed in a transfer line, as in a plug-flow or riser reactor, all three terms implying the type of operation shown in Figure 28.

For the plug-flow reactor Equation (7) can be expressed as sulfur concentration change with the distance

$$-\frac{dS}{dx} = k\ [S]^a\ [H_2O]^b \tag{35}$$

Due to large excess of steam, it can be assumed that $[H_2O] \neq f(x)$, and after separation of variables, integration of Equation (35) is possible.

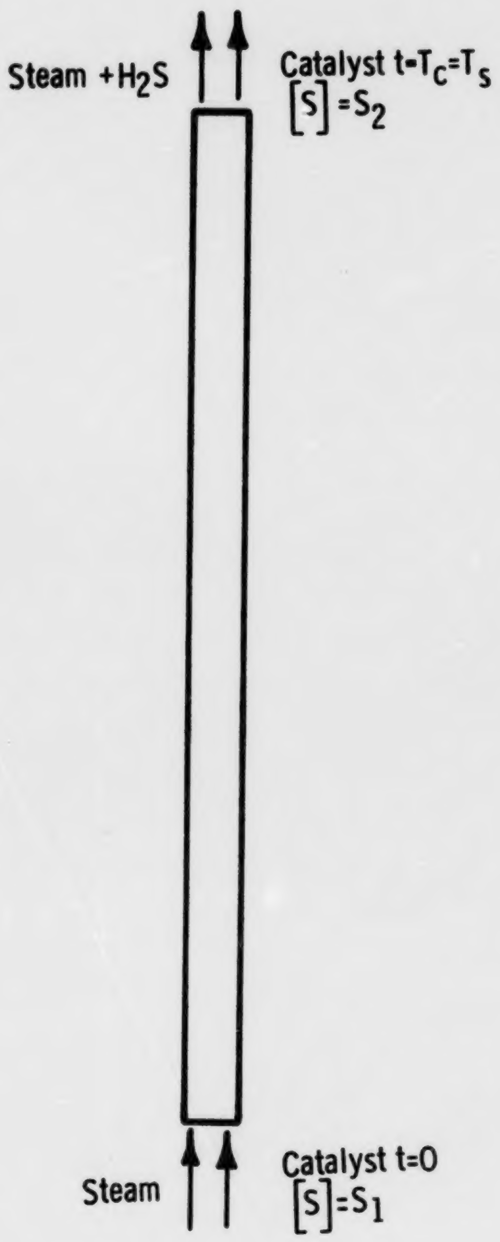
Steam +H_2S
Catalyst $t=T_C=T_S$
$[S]=S_2$
Steam
Catalyst $t=0$
$[S]=S_1$

Figure 28. Schematic of plug flow steam stripper

$$\frac{dS}{[S]^a} = -k[H_2O]^b \; dx \tag{36}$$

$$\frac{1}{1-a}\left[(S_2)^{(1-a)} - (S_1)^{(1-a)}\right] = -k[H_2O]^b \; L \tag{37}$$

for $a + 1$, and

$$\ln (S_2) - \ln (S_1) = -k[H_2O]^b \; L \tag{38}$$

for $a = 1$

where L = length of the plug reactor

Since $T_c = T_s$, the reaction length can also be expressed in terms of time as follows:

$$L = v \times T_c = v \times T_s \tag{39}$$

where v = velocity of steam-catalyst mixture through the reactor

Using Equations (16) and (31) the ratio of T_c/T_s can be determined:

$$\frac{T_c}{T_s} = \frac{\rho_c \overline{M}_s}{C \; \rho_s} \tag{40}$$

from which

$$T_c = \frac{\overline{M}_s}{C} \, T_s \, \frac{\rho_c}{\rho_s} \tag{41}$$

Expressing steam-to-catalyst ratio by steam stripping rate we can write

$$\frac{W_s}{C} = \frac{SSR}{100} \tag{42}$$

Now by substituting Equations (39), (41), and (42) into Equations (37) and (38) we obtain

$$\frac{1}{1-a}\left[(S_2)^{(1-a)}-(S_1)^{(1-a)}\right] = -k\,[H_2O]^b\ v\ T_s\ \frac{\rho_c}{\rho_s}\,\frac{(SSR)}{100} \tag{43}$$

for $a \neq 1$, and

$$\ln\,(S_2) - \ln\,(S_1) = -k[H_2O]^b\ v\ T_s\ \frac{\rho_c}{\rho_s}\,\frac{(SSR)}{100} \tag{44}$$

for $a = 1$.

Applying the simplification assumptions of $a = 1$ and $b = 1$, cancelling ρ_s with $[H_2O]$, and including $\rho_c/100$ into k, we obtain

$$\ln\frac{S_2}{S_1} = k\ v\ T_s\ (SSR) \tag{45}$$

Using Equation (12) we can write

$$\ln\,(1-X) = -k\ v\ T_s\ (SSR) \tag{46}$$

or

$$X = 1 - e^{-k\ v\ T_s\ (SSR)} \tag{47}$$

5.1.3.3 Counter-Current Stagewise Contacting -

The design of a counter-current, stagewise catalyst steam stripper is depicted schematically in Figure 29. For each of the stages in the contactor, the behavior of the fluidized bed is the same as in the continuous fluidized bed stripper. In this design, catalyst is flowing downward by gravity from stage to stage with each stage performing as an equal-size fluidized bed. Each stage uses the off-gases from the next lower stage for fluidization.

It is assumed in this model that the sulfur concentration in the vapor phase does not affect the sulfur removal efficiency. The sulfur material balance can be determined by considering the desorption of sulfur from the spent catalyst while disregarding the re-adsorption of sulfur by the catalyst. This assumption can only be verified by experimentation. Our experiments have supported the fact that the sulfur removal is controlled kinetically rather than by equilibrium. Consequently it is justified to assume that re-adsorption has a minimal effect. With this in mind, the counter-current reactor becomes a backmix reactor. Each stage is equal in size and can then be described as a continuous fluidized bed reactor and the following relationships for each stage of the contactor exist:

$$S_o - S_1 - k\, S_1{}^a\, [H_2O]^b\, T_{c1} = 0 \qquad (48)$$

$$\frac{S_o - S_1}{S_1{}^a} = k\, [H_2O]^b\, T_{c1} \qquad (49)$$

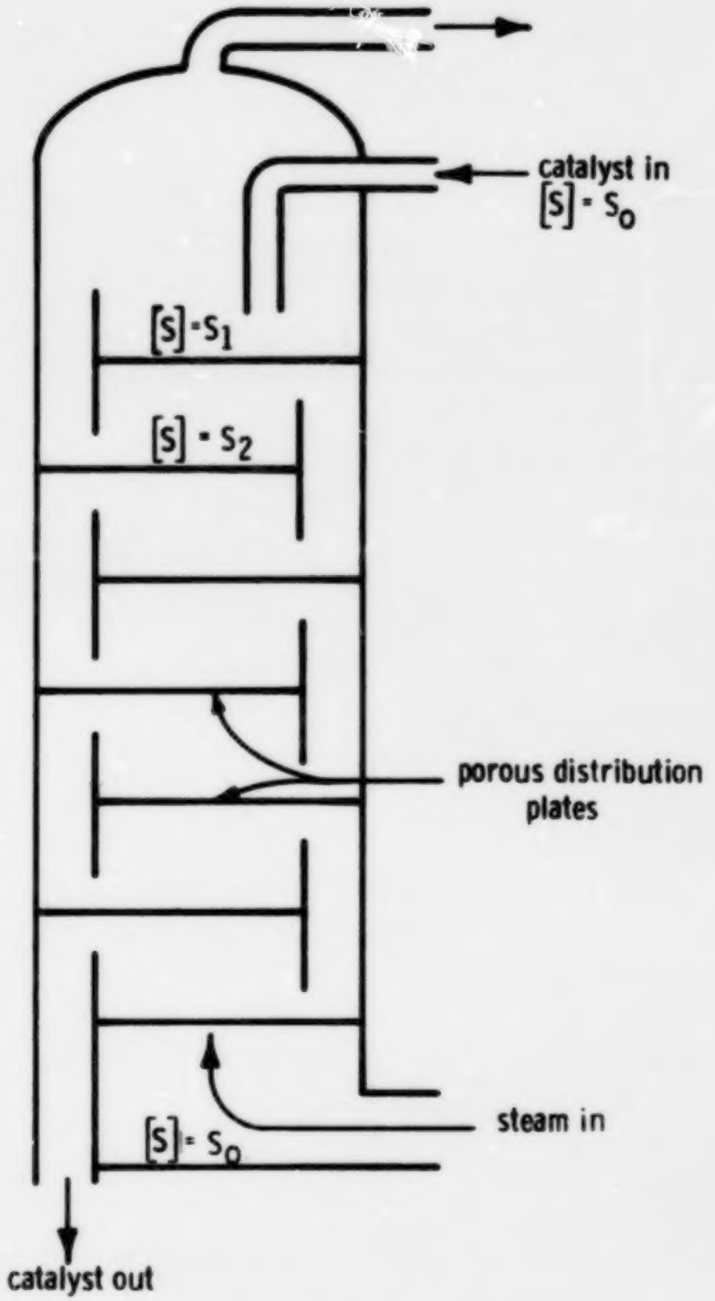
catalyst in
[S] = S_0
[S] = S_1
[S] = S_2
porous distribution plates
steam in
[S] = S_0
catalyst out

Figure 29. Counter-current, stagewise contactor

$$\frac{S_1 - S_2}{S_2{}^a} = k\ [H_2O]^b\ T_{c2} \tag{50}$$

$$\vdots$$

$$\frac{S_{n-1} - S_n}{S_n^a} = k\ [H_2O]^b\ T_{cn} \tag{51}$$

where n = number of reactor stages

Assuming a=1, and applying Equation (12) in the form

$$X_n = \frac{S_{n-1} - S_n}{S_{n-1}} \tag{52}$$

the set of Equations (49) through (51) will become

$$\frac{S_1}{S_o} = 1 - X_1 = \frac{1}{1 + k\ [H_2O]^b\ T_{c1}} \tag{53}$$

$$\frac{S_2}{S_1} = 1 - X_2 = \frac{1}{1 + k\ [H_2O]^b\ T_{c2}} \tag{54}$$

$$\vdots$$

$$\frac{S_n}{S_{n-1}} = 1 - X_n = \frac{1}{1 + k\ [H_2O]^b\ T_{cn}} \tag{55}$$

Multiplying Equations (53) through (55) and assuming

$$T_{c1} = T_{c2} = \dots\dots = T_{cn} \tag{56}$$

we obtain

$$\frac{S_n}{S_o} = \frac{1}{\left[1 + k\ [H_2O]^b\ T_{cn}\right]^n} \qquad (57)$$

Defining sulfur removal efficiency for the whole reactor as

$$X = \frac{S_o - S_n}{S_o} = 1 - \frac{S_n}{S_o} \qquad (58)$$

and catalyst contact time in the reactor as

$$T_c = n\ T_{cn} \qquad (59)$$

we can modify Equation (57) as follows:

$$1-X = \frac{1}{\left[1 + k\ [H_2O]^b\ T_c/n\right]^n} \qquad (60)$$

or

$$X = 1 - \left(\frac{1}{1 + k\ [H_2O]^b\ T_c/n}\right)^n \qquad (61)$$

Using Equations (16), (31), and (32), assuming b=1, applying assumptions from page 109, and including n in k, Equation (61) will yield

$$X = 1 - \left(\frac{1}{1 + k\ T_s(SSR)}\right)^n \qquad (62)$$

5.2 CONDENSER DESIGN

The catalyst stripper effluent steam may contain H_2S, NH_3, and hydrocarbons (see Tables 20 through 21). This steam is condensed to recover heat and to achieve a separation of H_2S from the water. A common shell and tube heat exchanger may be used with the stripper off-gas being passed through the tube side of the heat exchanger. Cooling water, on the heat exchanger shell side, is used as process water for subsequent steam stripping. The stripping steam is cooled from 811 K (1000°F), condensed, and subcooled while the process water is vaporized to produce saturated steam at 9.63×10^5 Pa (125 psig).

In order to minimize corrosion in the condenser, the materials of construction should be at least a low grade stainless steel, such as 5% Cr plus 1/2% Mo alloy.* However, the current trend is to more expensive stainless steels such as types 321 and 347 after stabilized annealing.**

The composition of the process water should meet or surpass the boiler feedwater specifications before it enters the steam superheater. These specifications are presented in Table 28.

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*Fontana, M. G., and N. D. Greene. Corrosion Engineering. New York, McGraw-Hill Book Co., 1967.

**Evans, F. L. Refiners Face Corrosion Facts. Hydrocarbon Processing. 53:109-112, April 1974.

Table 28. RECOMMENDED LIMITS OF SOLIDS IN BOILER FEEDWATER[a]

Drum pressure[b]	Below 600 psi	600 to 1000 psi	1000 to 2000 psi	Over 2000 psi
Total solids, ppm			0.15	0.05
Total hardness as ppm $CaCO_3$	0	0	0	0
Iron, ppm	0.1	0.05	0.01	0.01
Copper, ppm	0.05	0.03	0.005	0.002
Oxygen, ppm	0.007	0.007	0.007	0.007
pH	8.0-9.5	8.0-9.5	8.5-9.5	8.5-9.5
Organic	0	0	0	0

[a] Steam/its generation and use. New York, Babcock and Wilcox, 1972

[b] $1 \text{ psi} = 6.895 \times 10^3$ Pa

In our cost analysis, the design of the condenser was based upon the normal heat exchanger design equation (Phase I final report, Appendix E)

$$Q = UA\Delta T_m \qquad (63)$$

where U was assumed to be 3975 W/m^2K (700 $Btu/ft^2 \cdot {}^\circ F \cdot hr$). The ΔT_m was calculated from the heat exchanger terminal temperatures according to the following example:

811 K (1000°F) = stripper off-gas inlet to heat exchanger
311 K (100°F) = stripper off-gas outlet from heat exchanger
300 K (80°F) = process water inlet
451 K (353°F) = 9.63×10^5 Pa (125 psig) steam, saturated

$$\Delta T_m = \frac{(811-451) - (311-300)}{\ln \dfrac{811-451}{311-300}} = 100.1 \text{ K} \qquad (64)$$

Based on the above assumptions and knowing the amount and conditions of steam used for catalyst steam stripping, the heat transfer area for the condenser can be calculated from Equation (63).

$$A/Q = \frac{1}{3975 \times 100.1} = 2.514 \times 10^6 \text{m}^2/\text{W} \text{ (7.92 sq ft/}10^6\text{Btu/hr)} \qquad (65)$$

5.3 ACIDIFIER/PHASE SEPARATOR DESIGN

5.3.1 Equilibrium Relationship

The design of the acidifier/phase separator system will be dictated by several factors. These include the hydrogen sulfide content of the steam stripper off-gas, the temperature, the system pressure, the hydrogen ion concentration (pH) of the condensate, and the allowable H_2S concentration of the effluent wastewater. The H_2S content of the stripper gas will be dictated by the sulfur content of the spent catalyst and the efficiency of the stripper. However, H_2S concentrations of up to 2.0×10^3 mole fraction (2000 vppm) can be expected in the stripper overhead vapors. The pH of the condensate and its sulfide content may also be affected by the ammonia content of the stripper off-gas. The allowable H_2S content of the phase separator condensate will be dictated by either federal, state, or local EPA standards for refinery effluents. However, a sulfide concentration 10^{-3}kg/m^3 will probably be the highest tolerable level.*

It should be noted that as long as the condensate containing dissolved and unreacted H_2S (g) is exposed to ambient air after leaving the condenser, the H_2S gas will diffuse out of the liquid and leave zero H_2S concentration level. The rate of this diffusion will depend on the H_2S concentration, the temperature of the condensate, and the effectiveness of liquid-air contacting downstream of the condenser. Consequently, the final residual sulfide concentration in the condensate effluent will be a function of the diffusion rate, the amount of mercaptans condensed, and the amount of compounds that can tie with H_2S and form sulfide salts.

Since essentially pure steam is used in the steam stripping, the compounds that can react with H_2S to form sulfides must be formed in the steam stripping process. Our experiments showed that only one such compound is formed, ammonia.

Thus, the total sulfide concentration [TSS] in the condenser water effluent may be expressed as follows

$$[TSS] = [H_2S] + [MSH] + [(NH_4)_2S] \tag{66}$$

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*Topical Law Reports, Pollution Control Guide. New York, Commerce Clearing House, Inc., Vol. 2, Part 419, pp. 9627-9627-19.

where $[H_2S]$ = dissolved hydrogen sulfide
$[MSH]$ = mercaptan sulfide
$[(NH_4)_2S]$ = ammonium sulfide

Data from our experiments indicated that practically no formation of mercaptans occurs. Hence, the second term of Equation (66) may be neglected.

$$[TSS] = [H_2S] + [(NH_4)_2S] \tag{67}$$

As shown in Table 22, we have observed ammonia formation in concentrations ranging from 3.4×10^{-4}-5.2×10^{-4}mole fraction (340 to 520 ppm).

Table 29 summarizes the first step dissociation constants for hydrogen sulfide in the temperature range between 5 and 60°C.* Assuming a temperature of 25°C we can calculate the relative contents of species produced from hydrogen sulfide as a function of pH, Figure 30. The second step dissociation constant at this temperature is 1.0×10^{-15}.* Both constants can be written as follows

$$K_1 = \frac{[H^+]\,[HS^-]}{[H_2S]} \tag{68}$$

$$K_2 = \frac{[H^+]\,[S^=]}{[HS^-]} \tag{69}$$

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* Gmelins Handbuch Der Anorganischen Chemie (Gmelins Handbook of Inorganic Chemistry), 9th Edition, Number 9, Sulfur, Section B1. Weinheim/Bergstrasse, Germany; Verlag Chemie, GMBH; 1953.

Table 29. IONIZATION CONSTANTS FOR THE H_2S-WATER SYSTEM AT VARIOUS TEMPERATURES

Temperature,[a] °C	$K_1 \times 10^7$
5	0.471
10	0.574
15	0.747
18	0.910
20	0.853
25	1.08
25	1.15
30	1.26
40	1.64
50	2.03
60	2.39

[a] $t_k = tc + 273.15$.

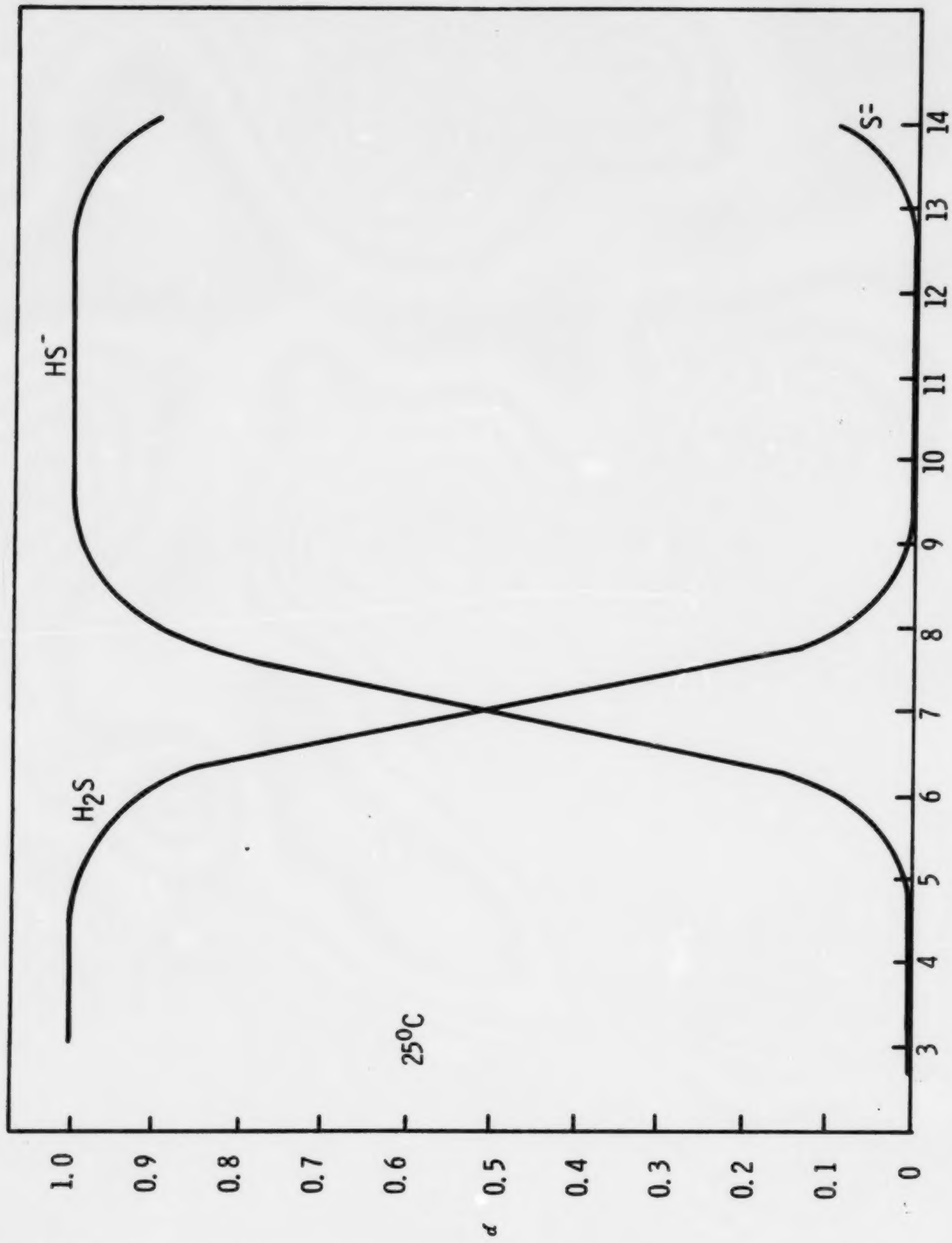

Figure 30. Distribution diagram for hydrogen sulfide (Note that [$S^=$] has appreciable concentration only in strongly basic solutions)

For our purposes, K_1 values of up to 100°C are needed. They were obtained by fitting the data presented to an empirical equation below. The equation shown also in Figure 31, predicts the ionization constant at any temperature in the range of 0 to 60°C with greater than 99.9% confidence.

$$K_1 = [(0.0356655)T + 0.2288326] \times 10^{-7} \qquad (70)$$

where T = temperature (°C)

In the absence of data for temperatures above 60°C, an extrapolation of data found in the literature was made. The degree of reliability of such an approach, however, should be verified by ascertaining experimental data at these temperatures since significant deviations from extrapolated values may occur. The above equation may be used to obtain appropriate K_1 values at various temperatures over the range of 0 to 199°C.

As indicated in Figure 30 the $[S^=]$ concentration becomes significant only at high pH ($pH \geq 13$) or in strongly basic solutions. In our case where there is an excess of hydrogen sulfide present the pH of the condensate will never approach high pH values and formation of $[S^=]$ may be neglected. Consequently, the species present in the condensate will include HS^- and H_2S. All ammonia will also be in the form of NH_4HS.

Increasing temperatures will allow higher ratios of ionized species at lower pH values. This change, however, may be considered insignificant since a 35°C increase will move the curves in Figure 30 by only 0.3 pH unit to the range of lower pH values.

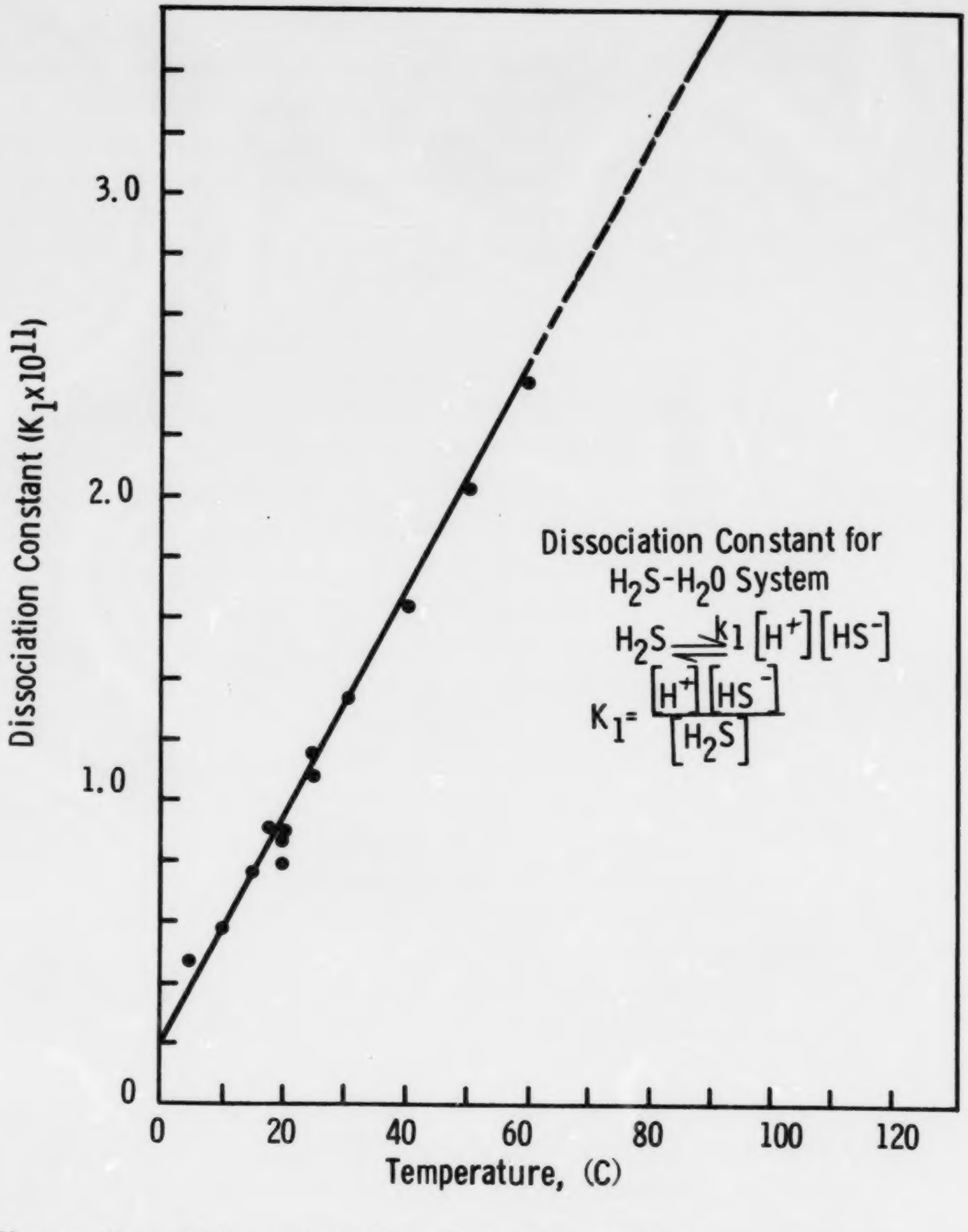

Figure 31. Effect of temperature upon ionization constant for H_2S

In case some hydrogen sulfide is present in the gaseous phase above the condensate, the concentration of total sulfur can be determined from Henry's law.

$$[H_2S]_g = H\,[H_2S]_{aq} \qquad (71)$$

where H = Henry's law constant for hydrogen sulfide (atm/mole fraction)

$[H_2S]_g$ = partial pressure of H_2S in gaseous phase (atm)

$[H_2S]_{aq}$ = mole fraction of hydrogen sulfide in solution

The values of Henry's law constant for temperatures between 0 and 100°C are presented in Table 30.

The condenser material balance can now be determined by means of Equation (71) and Equation (72) if an assumption of steam condensation ratio is made.

$$V\,(1-y)\,\frac{[H_2S]_g}{\pi} + Vy\,[H_2S]_{aq} = Vx \qquad (72)$$

where V = volume of steam (moles)

y = fraction of steam condensed

x = mole fraction of H_2S in steam entering the condenser

π = condenser pressure (atm)

Volume of steam occurs in each term of Equation (72) and may be cancelled. $[H_2S]_g$ may be expressed from Henry's law, Equation (71), and we obtain

$$(1-y)\,\frac{H\,\left[H_2S\right]_{aq}}{\pi} + y\,[H_2S]_{aq} = x \qquad (73)$$

TABLE 30. HENRY'S LAW CONSTANT FOR H_2S VERSUS TEMPERATURE[a]

Temperature, °C	Henry's Law Constant[b] atm/mole fr. H_2S in sol.
0	20,300
10	36,700
20	48,300
30	60,900
40	74,500
50	88,400
60	103,000
70	119,000
80	135,000
90	144,000
100	148,000

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[a] Perry, J.H., Chemical Engineers' Handbook, 4th Edition. New York, McGraw-Hill Book Co., 1963.

[b] 1atm = 1.01325×10^5 Pa.

Knowing the concentration of H_2S in the stripper effluent $\underline{x}$, and Henry's law constant at various temperatures and degrees of steam condensation y at the operating temperature of the condenser, we can determine condenser pressure from steam tables and subsequently calculate $[H_2S]_{aq}$. Substituting $[H_2S]_{aq}$ in Equation (71) with proper H we can determine concentration of H_2S in vapor phase.

Results obtained for a stream containing 2.0×10^{-3} mole fraction (2000 ppm) H_2S at three different temperatures and various fractions of steam condensation are presented in Figure 32.

As can be seen in Figure 32, the condensation of stripping steam containing 2.0×10^{-3} mole fraction (2000 ppm) H_2S may produce a large range of concentrations in vapor and liquid phase depending on the operating conditions maintained in the condenser. Specifically, 99% condensation at temperatures lower than about 60°C will not result in liquid concentrations that would exceed 1.0×10^{-3}kg/m^3 (1 ppm). This is valid, of course, only if no ammonia is present. Condensation at higher temperatures or higher condensation ratios would produce solution with H_2S concentrations higher than 1.0×10^{-3}kg/m^3 (1 ppm).

As pointed out earlier, once the solution of H_2S is exposed to the gas phase containing no H_2S, the diffusion of H_2S from liquid phase to gaseous phase would occur until a new equilibrium was reached. If the liquid were exposed to ambient air the H_2S concentration would eventually go to zero.

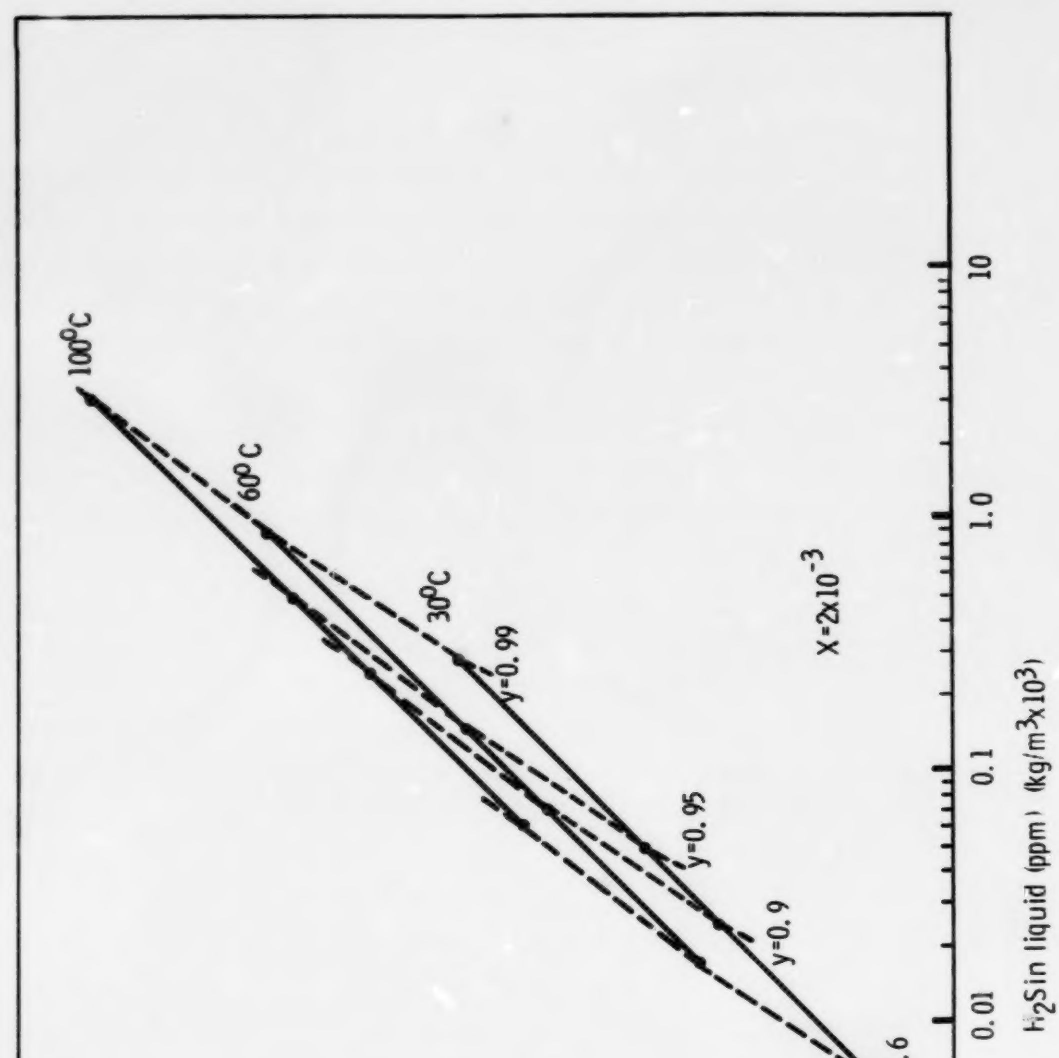

Figure 32. H_2S concentration in vapor and liquid stream

Analyzing the hydrogen sulfide dissociation constant, Equation (68) or Figure 30, we see that an increase of hydrogen ion concentration would shift the equilibrium to non-dissociated H_2S. Since only non-dissociated H_2S can diffuse out of solution this would increase the driving force of the H_2S diffusion to gaseous phase and enhance the rate of H_2S depletion. A pH value of about 5 would convert essentially all dissolved sulfide species to non-dissociated H_2S and result in the maximum rate of H_2S depletion.

Essentially the same principles apply in the presence of ammonia. Some of the acid, however, will react with the ammonia and increase the acid consumption.

Applying the above principles, we may conclude that an increase of hydrogen ion concentration by injection of an acid into the condenser would further lower the H_2S concentrations in the condensate and increase H_2S vapor contents. pH values around 5 would cause the H_2S to stay in vapor phase and prevent it from going into the condensate. Thus, very rich hydrogen sulfide vapor stream and sulfide free condensate would result.

In the design of the steam condenser a trade-off will have to be evaluated between efficiency of the H_2S recovery by treating the condensate, or handling acidic streams at elevated temperatures that may range from 300 to 373K (80-212°F). In one case, the acid treatment may be done in a condensate tank. However, the depletion of H_2S from the bulk of the condensate will be diffusion controlled. In the second case, the condenser interior has to be acid resistant at condenser operating conditions and no diffusion factors apply since the H_2S will never enter the condensed phase.

5.3.2 Combined Condenser/Acidifier/Phase Separator Design (Alternative Sour Water Treatment System)

In the process outlined in the Phase I final report (page 63), condensation, phase separation, acidification, and clarification occur in four separate vessels. Retrofit of such a system to existing FCC units may be difficult because of severe space limitations which exist at petroleum refineries. A system which accomplishes all four processing steps plus separation of oil from the water in an integrated space-saving unit is desirable. One approach to design of such a system is proposed and presented in Figure 33. Its operation is described below.

Acidification of the condensate is achieved by injecting sulfuric acid into the stripper off-gas vapor through a spray nozzle. Two or more nozzles should be available to prevent shut-down precipitated by possible corrosion of the nozzle. The H_2SO_4 injected into the 811 K (1000°F) stream will decompose and form sulfur trioxide. To ensure homogeneous distribution of H_2SO_4 and SO_3 in the vapor and the condensate, the vapor stream should be properly mixed. A motionless mixer is visualized to fulfill this function. Vapors containing H_2S, NH_3, SO_3, hydrocarbons and catalyst fines are condensed in a vertical floating head heat exchanger. The H_2SO_4 injected into the vapor stream is used to control the pH of the condensate.

A vapor-liquid separating cone is located in the bottom of the condenser. Liquids formed in the condenser are sent to a liquid surge drum. The vapors pass through a demister pad and are sent to an existing refinery Claus unit for sulfur recovery.

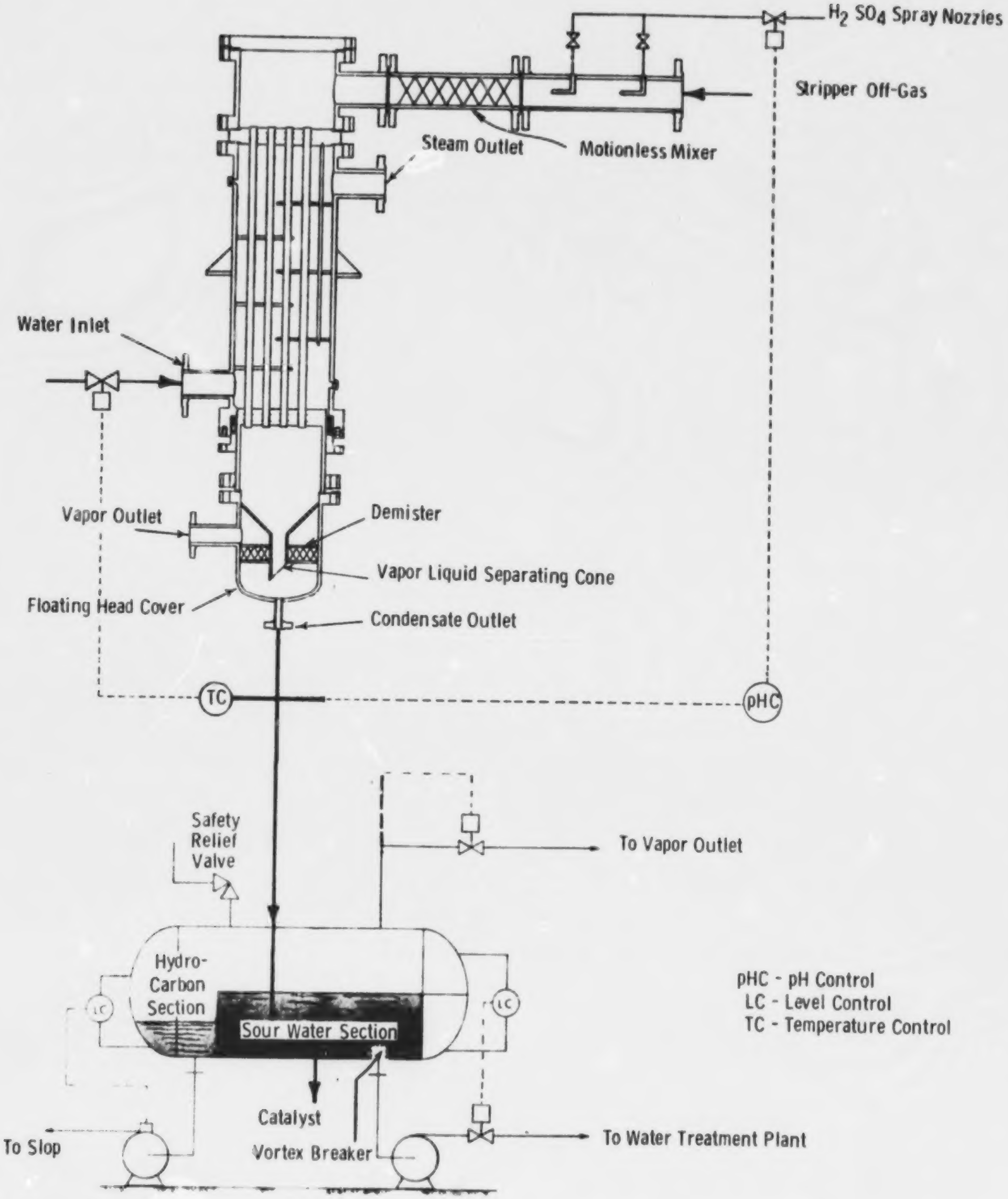
H2 SO4 Spray Nozzles
Stripper Off-Gas
Steam Outlet
Motionless Mixer
Water Inlet
Vapor Outlet
Demister
Vapor Liquid Separating Cone
Floating Head Cover
Condensate Outlet
TC
pHC
Safety Relief Valve
To Vapor Outlet
Hydro-Carbon Section
Sour Water Section
LC
LC
pHC - pH Control
LC - Level Control
TC - Temperature Control
Catalyst
Vortex Breaker
To Slop
To Water Treatment Plant

Figure 33. Combined condenser, acidifier, phase separator, clarifier, and scum oil removal system

The liquid surge drum is used to achieve several processing steps simultaneously - oil and water separation, catalyst fines removal (clarification), and residual H_2S removal.

Water leaving the liquid surge drum is to be sent to the refinery wastewater treatment for disposal or recycle.

6. ECONOMICS

In the Phase I final report we proposed two possible conceptual methods of applying steam stripping to existing FCC units:

1. An increase of the present steam stripping rates in existing equipment to the levels needed to achieve an adequate SO_x reduction in the generator flue gas.

2. The use of a secondary open (add-on) stripping system in which the spent catalyst is removed from the existing equipment and transferred to a secondary stripper. Sulfur free catalyst is then returned to the regenerator.

Both these methods were evaluated in Section 5.3 of the Phase I final report. The first method was found economically infeasible due to limited size of existing stripper, reactor, overhead vapor line and FCC fractionator, and the inability of this equipment to handle increased flow rates resulting from supplemental stripping steam.

The second method was also evaluated in detail and several options were proposed for SO_x emission control by steam stripping. Option 1 was believed to be the most unfavorable economically and was analyzed further in two cases, the worst and the typical case. Both cases represented the cost of overall system presented in Figure 13, p. 63, Phase I final report including the waste water treatment facilities.

A detailed description and analysis of the Option 1 and definitions of the worst and typical cases were presented in Section 5.3.3 and Appendix E of the same report.

In evaluating the worst and the typical cases of Option 1 several assumptions were made and are repeated below:

	Worst Case	Typical Case
Catalyst-to-oil ratio	12	6
Catalyst attrition rate, kg/m^3	0.57	0.29
lb/barrel	0.2	0.1
Steam stripping rate, kg H_2O/100 kg of catalyst	4	4

The data generated during the laboratory development program revealed that a variety of steam stripping rates may be required for different catalysts. Specifically, rates as high as 100kg/100kg of catalyst were required to obtain 2×10^{-4} mole fraction (200 vppm) of sulfur oxides in the FCC regenerator off-gas.

Change of steam stripping rate will require a change in the equipment capacity for each processing step and result in different process economics. Consequently, we have expanded the economic evaluations performed in Phase I to higher stripping rates such as 6, 12, 40, and 100kg of steam per 100kg of catalyst. All the assumptions used in preparation of these evaluations were the same as those applied in the Phase I report, which makes the results of both reports (Phase I and Phase II) comparable. Also, by using the results from the estimates prepared in Phase I for the steam stripping rate of 4kg/100kg of catalyst, our overall range of steam stripping rate is from 4 to 100kg/100/kg catalyst. Using this range of steam stripping rates allowed presentation of the capital investment and operating costs as a function of both steam stripping rate and refinery size.

As in Phase I, estimates were prepared for both the typical and worst cases. The results of these evaluations are summarized in Table 31 and Figures 34 through 39. Detailed cost estimates and calculations are presented in Appendix B.

It should be noted that one of the very important assumptions used in the economic analysis was the steam superficial velocity in the stripper. As indicated previously [Equation (3)], the sulfur removal efficiency appears to be a function of the product steam residence time in the stripper and steam stripping rate. The steam residence time is inversely proportional to steam superficial linear velocity which is a function of steam stripper design. With current stripper designs, the steam residence time can be varied over a wide range of values. Specifically, superficial linear velocity in the stripper can vary from $1.52x10^{-2}$7.62m/s (0.05 to 25 ft/s) a factor of 500.

Equation (3) suggests that an increase in steam residence time in the stripper can substantially reduce steam stripping rate with no change in sulfur removal efficiency. This can be easily done by decreasing the stripping steam superficial velocity. It will be shown that the reduction in steam stripping rate will have an influence on the economics of the steam stripping concept because the steam superheater capacity will be reduced, the capacities and size of equipment downstream the stripper will be reduced, and condensate treatment will be minimized.

Table 31. CAPITAL AND OPERATING COST SUMMARY FOR STEAM STRIPPING CONCEPT

FCC Unit Capacity

Operating Conditions	10,000 bpsd		50,000 pbsd		150,000 pbsd	
	Capital Investment Cost ($x10⁻⁶)	Operating Cost[c] (¢/bbl)	Capital Investment Cost ($x10⁻⁶)	Operating Cost[c] (¢/bbl)	Capital Investment Cost ($x10⁻⁶)	Operating Cost[c] (¢/bbl)
Typical stripping operation						
S/C^b = 4	0.45	13.84	1.30	7.20	2.82	5.96
S/C^b = 6	0.58	15.14	1.71	8.34	3.59	6.83
S/C^b = 12	0.92	19.09	2.73	11.51	5.75	9.64
S/C^b = 40	2.08	35.50	6.17	24.75	13.05	21.63
S/C^b = 100	3.86	65.28	11.52	51.21	24.46	45.75
Worst stripping operation						
S/C^b = 4	0.72	19.82	2.08	13.73	4.51	10.99
S/C^b = 6	0.92	22.09	2.73	14.27	5.76	12.64
S/C^b = 12	1.47	29.34	4.37	20.43	9.21	17.84
S/C^b = 40	3.32	57.00	9.90	45.20	19.47	40.77
S/C^b = 100	6.17	114.7	18.51	94.50	[illegible].45	87.04

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[a] 1 bpsd = 1.84x10⁻⁶ m^3/s

[b] S/C = Steam to catalyst ratio

[c] 1 bbl = 0.15899 m^3

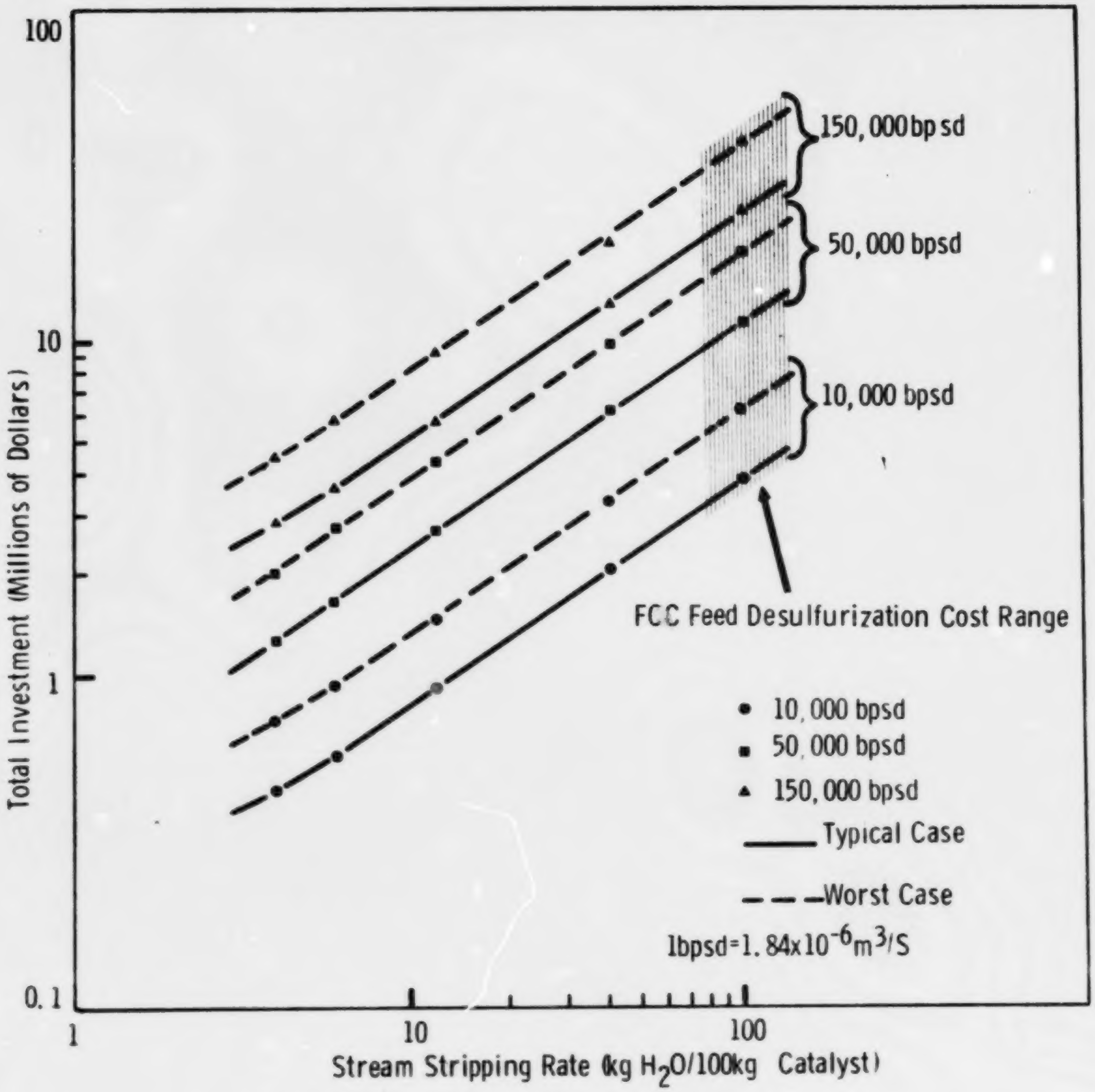


Figure 34. FCC catalyst steam stripping, total investment cost

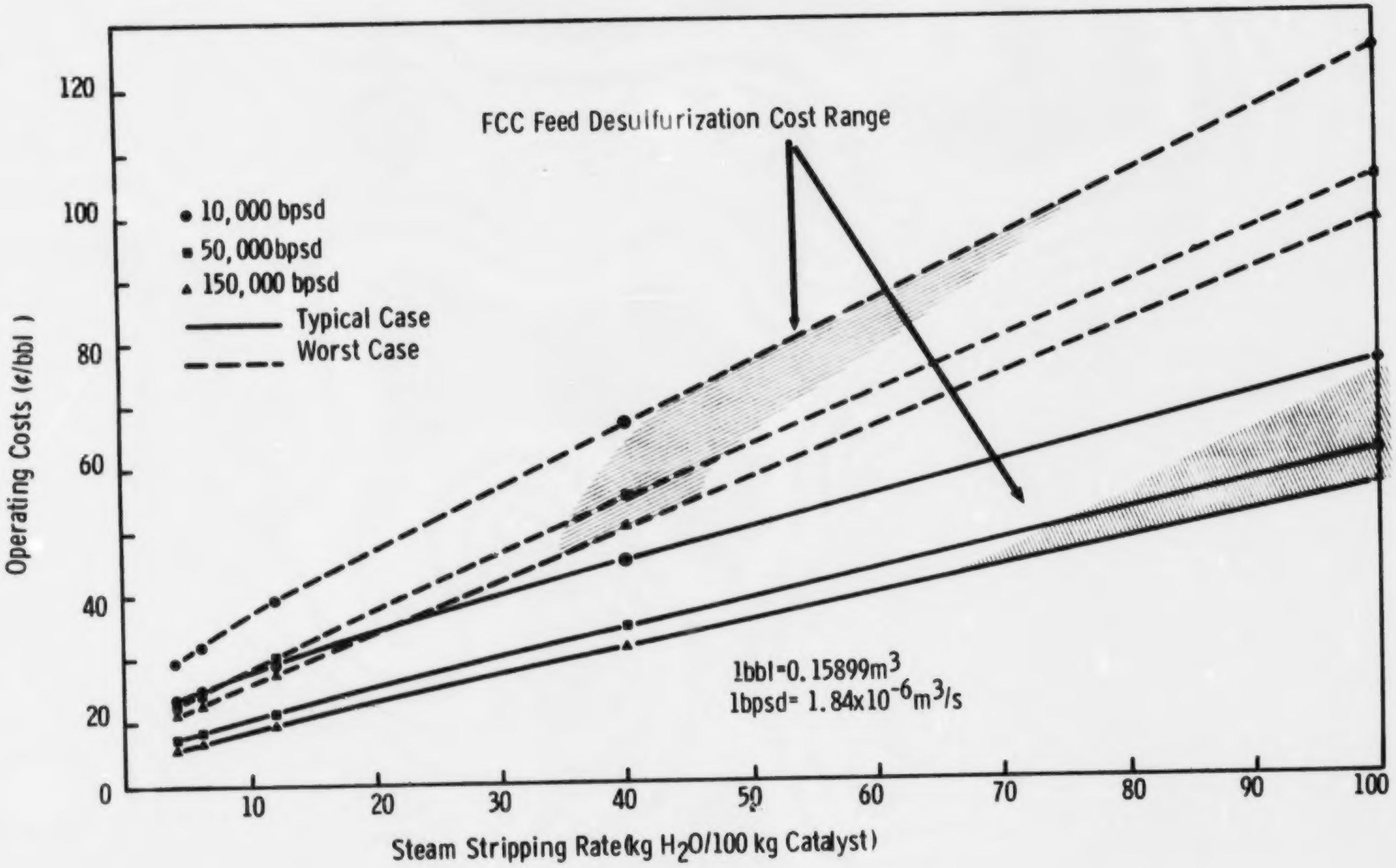


Figure 35. Summary of operating costs

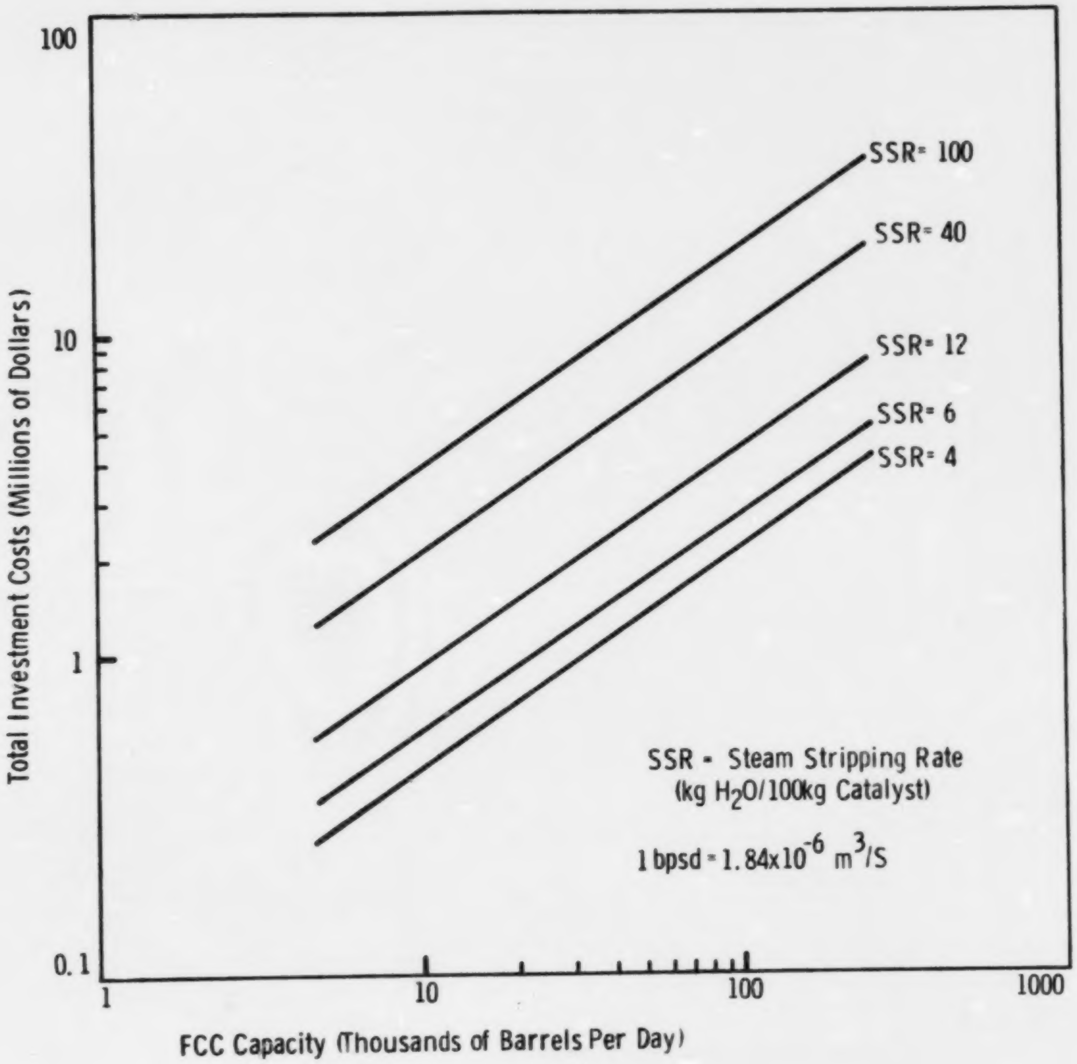


Figure 36. Summary of total investment costs, typical case

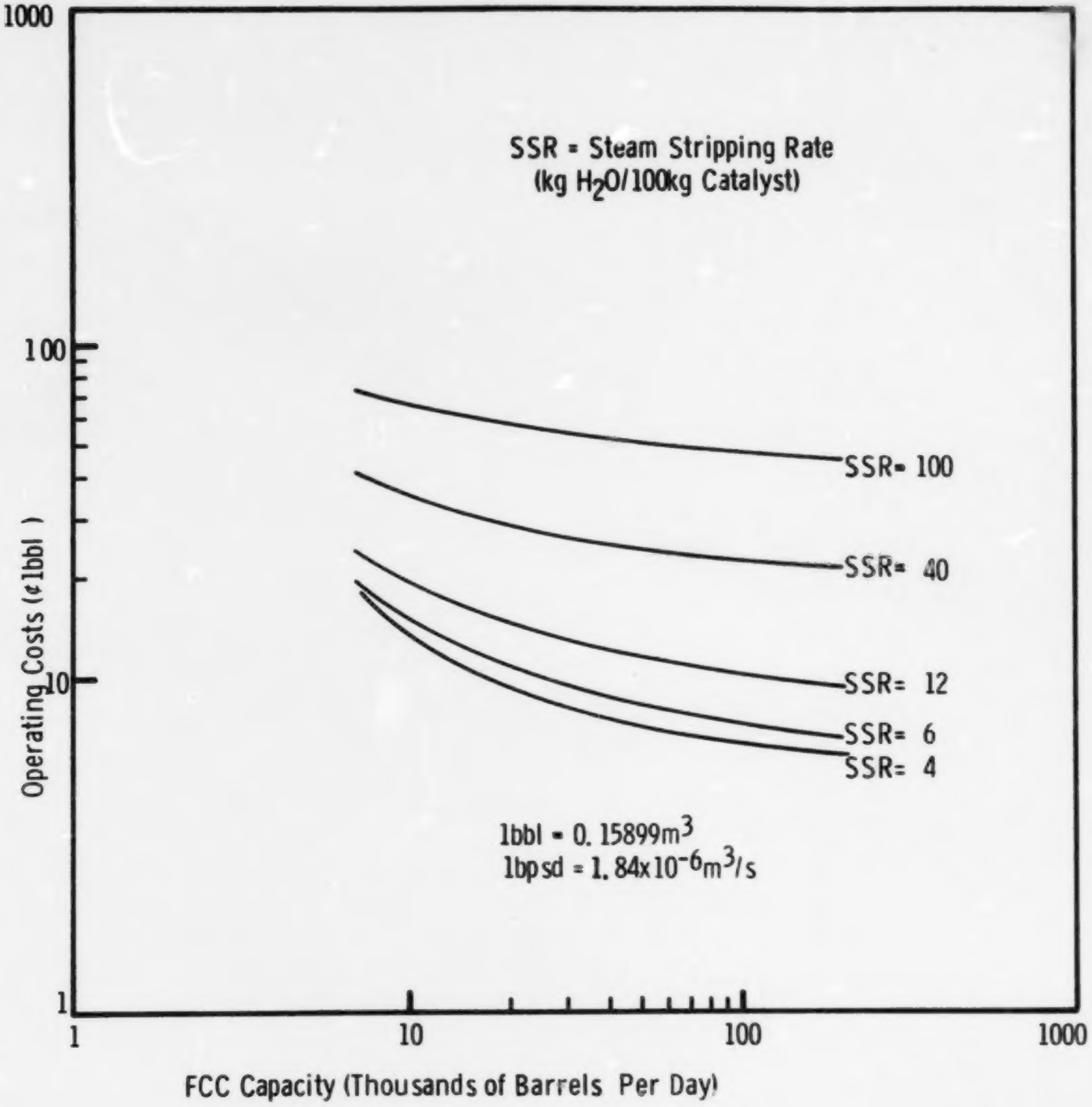


Figure 37. Summary of operating costs, typical case

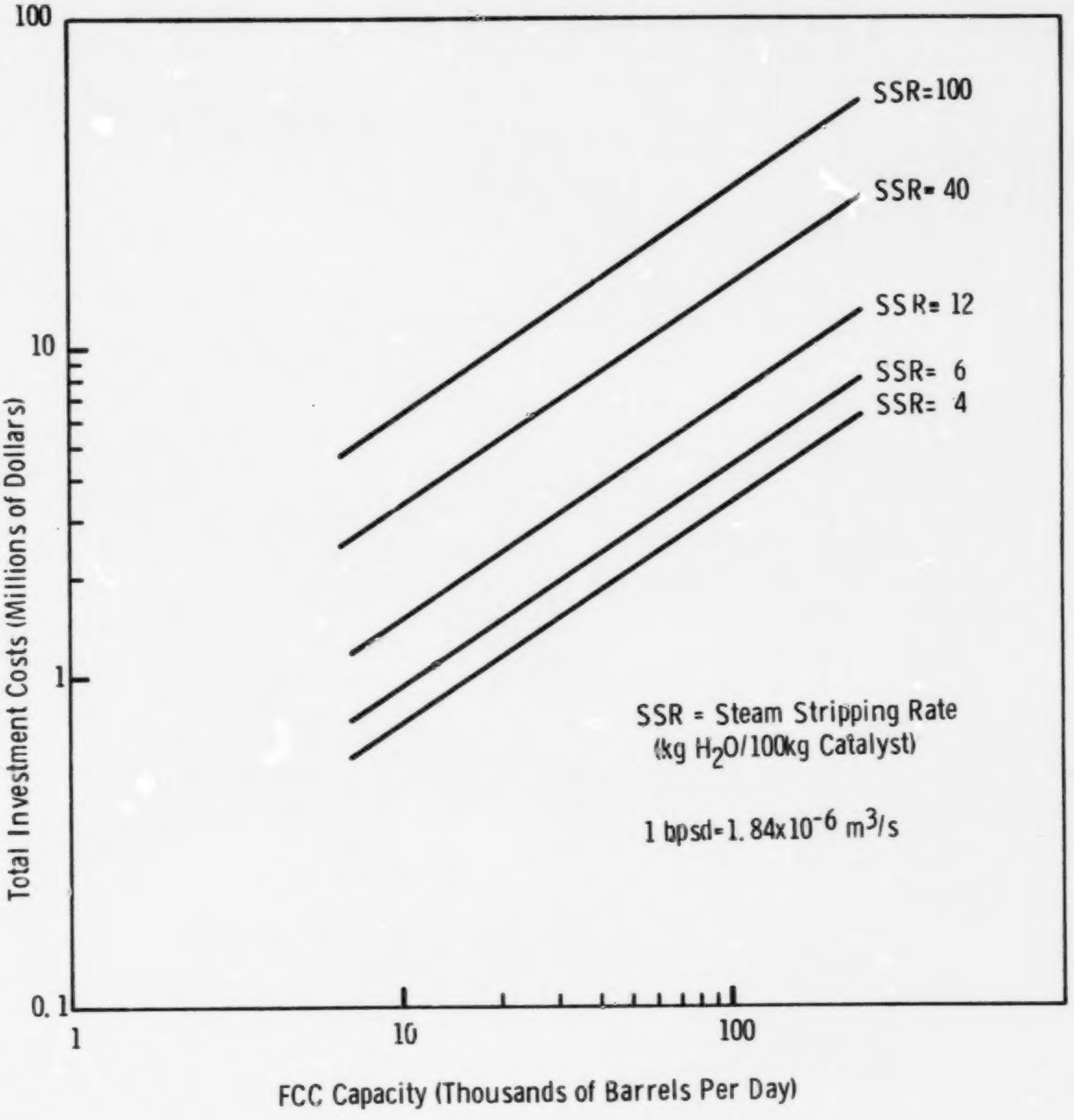


Figure 38. Summary of total investment costs, worst case

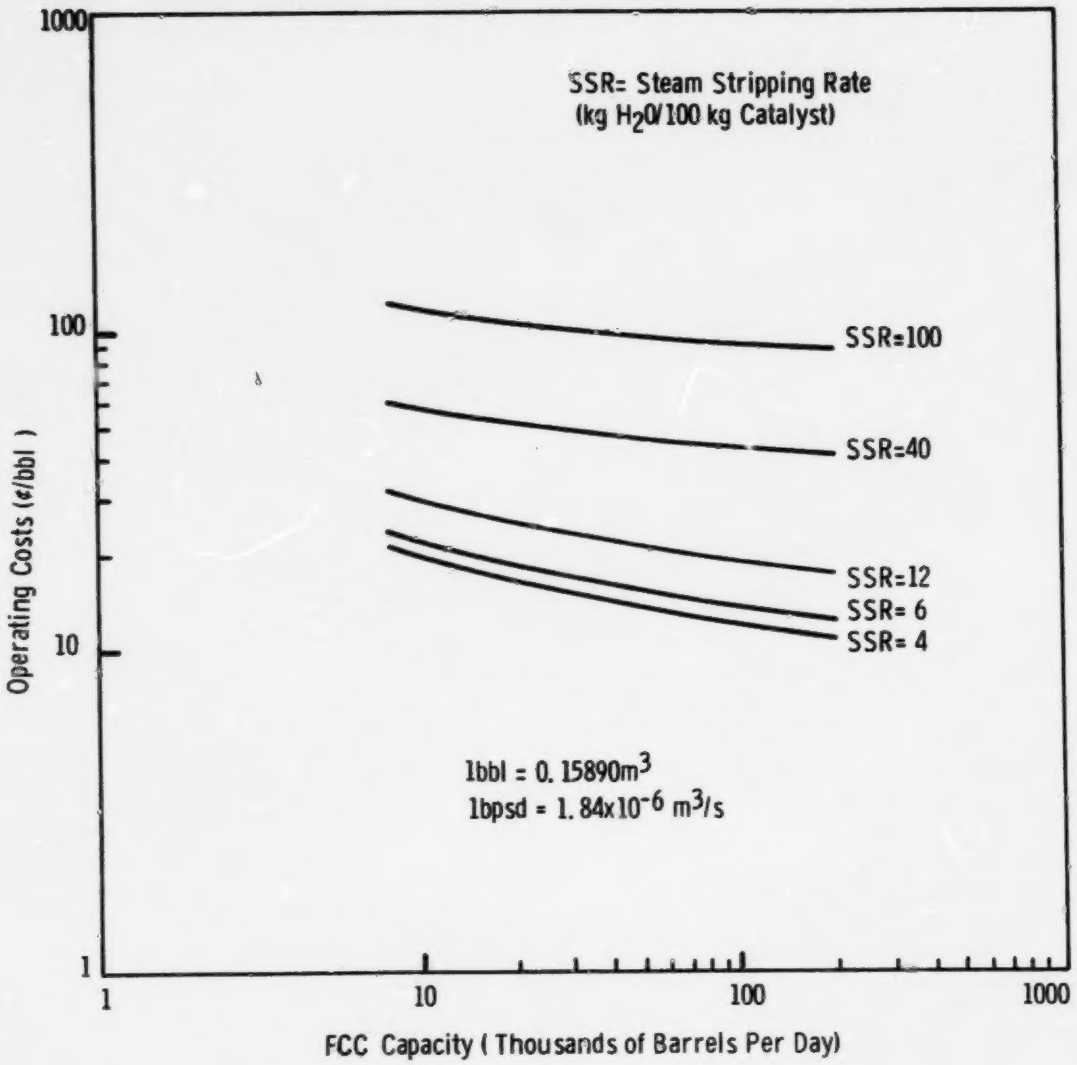


Figure 39. Summary of operating costs, worst case

In our cost estimate we have assumed a steam superficial velocity in the stripper of 0.61m/s (2 ft/s). In the following paragraphs we will discuss the effect of this velocity on the overall economics of the steam stripping concept. We will use an example of decreasing the velocity from 0.61 to 0.30m/s (2 to 1 ft/s). Essentially the same technique can be used to obtain costs of steam stripping at any velocity.

The reduction of steam superficial velocity from 0.61 to 0.30m/s (2 to 1 ft/s) will decrease the steam stripping rate by 50%. Since the size of most of the equipment used in the process is based upon steam capacity (all equipment except the stripper, which is based on catalyst residence time), the investment and operating costs will also decrease. The investment cost will decrease according to the 0.67 power of steam capacity. The raw materials and utilities costs will be cut in half.

Using the data in Table 31, the investment and operating costs for $9.20\times10^{-2}m^3/s$ (50,000 barrels/day) FCC unit capacity with steam stripping rate of 6 and 12kg of steam per 100kg of catalyst, and other typical stripping conditions, we can summarize

Steam stripping rate, kg H_2O/100 kg catalyst	Investment cost, millions $	Operating costs,[a] ¢/bbl
6	1.71	8.34
12	2.73	11.51

---

[a] 1 bbl = 0.15899 m^3.

Reduction of the superficial linear velocity in the stripper from 0.61 to 0.30m/s (2 to 1 ft/s) for the steam stripping rate of 12kg/100kg would decrease the investment and operating costs to $1,746,200 and 9.54 ¢/bbl, respectively. We can compare the costs for stripping velocities of 0.30m/s and 0.61m/s, both at a steam stripping rate of 6:

Stripping velocity		Catalyst residence time s	Total investment cost millions $	Operating costs,[a] (¢/bbl)
ft/sec	m/s			
2	0.61	104	1.71	8.34
1	0.30	209	1.75	8.33

---

[a] 1 bbl = 0.15899 m³.

As we can see, the costs for identical steam stripping rates and different stripping velocities are very well comparable. This means that the costs for any stripping velocity can be determined by proportionally reducing the steam stripping rate and obtaining the actual figures from cost data determined for stripping velocities of 0.61 m/s (2 ft/s). This technique should not result in an error larger than 5%.

The above example demonstrated that the superficial linear velocity is a very important factor in the economics of the steam stripping concept. Since the stripper velocities can vary over a range by a factor of 500, the economics of the steam stripping concept may also vary over a broad range and has to be carefully evaluated.

The economic analyses presented in this report do not include the additional economic benefits of steam stripping. The costs include only items incurred in performing the steam stripping operation. No by-product credit is taken for increased sulfur production, increased hydrocarbon recovery, and heat recovery from stripping steam upon its condensation. Each of these three factors would further improve the economics of the steam stripping concept.

APPENDIX A

SPENT FCC CATALYST STEAM STRIPPING DATA SHEET

I. EXPERIMENT NUMBER - ______________________________

II. DATE - ______________________________

III. EXPERIMENT PERFORMED BY - ______________________________

IV. SPENT FCC CATALYST IDENTIFICATION

A. COMPANY - ______________________________

B. REFINERY LOCATION - ______________________________

C. FCC UNIT - ______________________________

D. CATALYST TYPE - ______________________________

E. CATALYST HISTORY - ______________________________

____

V. PURPOSE OF THIS EXPERIMENT - ______________________________

____

VI. VARIABLES TO BE TESTED - ______________________________

____

VII. CATALYST ANALYSIS

A. C_{sc} = COKE ON SPENT CATALYST

C_{sc} = __________% BY WEIGHT

B. S_c = SULFUR CONTENT OF COKE

S_c = __________% BY WEIGHT

VIII. EXPERIMENTAL DATA

A. CATALYST CHARGING DATA

1. CAT_r = CATALYST CHARGED TO REACTOR

   CAT_r = __________GRAMS

2. S_r = S [illegible] ARGED TO REACTOR

   S_r = [illegible] $(C_{sc}) \times (S_c) \times (0.1)$

   S_r = __________MILLIGRAMS

B. STEAM STRIPPING OF SPENT FCC CATALYST

1. T_{cb} = CATALYST BED TEMPERATURE

   T_{cb} = __________

2. WATER FLOW RATE SETTING = __________

3. $\overline{W}_{H_2O}$ = WATER FLOW RATE

   $\overline{W}_{H_2O}$ = __________MILLILITERS/MIN

4. T_{ss} = SUPERHEATED STEAM TEMPERATURE

   T_{ss} = __________

5. P_s = STRIPPER PRESSURE

   P_s = __________psig

6. ρ_{ss} = SUPERHEATED STEAM DENSITY

   ρ_{ss} = __________lb/ft^3 (FROM STEAM TABLES)

7. V_s = CALCULATED STRIPPER LINEAR VELOCITY

   $V_s = (0.00263) \times (\overline{W}_{H_2O}) \times (\frac{1}{\rho_{ss}})$

   V_s = __________ft/sec

8. ΔP_s = STRIPPER PRESSURE DROP

   ΔP_s = __________INCHES OF WATER

9. ρ_{cat} = STATIC CATALYST BED DENSITY

   ρ_{cat} = __________lb/ft^3

10. F_{ex} = FLUID BED VOLUME EXPANSION FACTOR

F_{ex} = ___________

11. $H_{fc} = (0.158) \times (CAT_r) \times (\frac{1}{\rho_{cat}}) \times F_{ex}$

H_{fc} = _________ft

12. VOL_{fb} = VOLUME OF FLUIDIZED BED

$VOL_{fb} = (0.0140) \times (H_{fc})$

VOL_{fb} = __________ft^3

13. VOL_{cat} = VOLUME OF CATALYST IN FLUID BED

$VOL_{cat} = (0.0022) \times (\frac{1}{\rho_{cat}}) \times (CAT_r)$

VOL_{cat} = _________ft^3

14. VOL_{void} = VOLUME OF FLUID BED NOT OCCUPIED BY CATALYST

$VOL_{void} = VOL_{fb} - VOL_{cat}$

VOL_{void} = __________ft^3

15. T_{sc} = STEAM/CATALYST CONTACT TIME

$$T_{sc} = 27{,}240 \frac{(\rho_{ss}) \times (VOL_{void})}{\overline{W}_{H_2O}}$$

T_{sc} = ___________SECONDS

16. T_s = TOTAL STRIPPING TIME

T_s = __________SEC.

17. W_{H_2O} = ESTIMATED WATER USAGE

$W_{H_2O} = (\frac{1}{60}) \times (T_s) \times (\overline{W}_{H_2O})$

W_{H_2O} = ___________MILLILITERS (GRAMS)

18. S/C = STEAM TO CATALYST RATIO

$$S/C = \frac{100 \; W_{H_2O}}{CAT_r}$$

S/C = __________ lb H_2O / 100 lb CATALYST

C. SYSTEM PURGING

1. NITROGEN FLOW RATE SETTING = ____________
2. NITROGEN FLOW RATE = ___________ l/min
3. PURGING TIME = ___________ min
4. PURGE VOLUME = ___________ l

D. COKE COMBUSTION

1. CATALYST BED TEMPERATURE - SEE ATTACHED SHEET
2. T_a = PREHEATED AIR TEMPERATURE

   T_a = __________
3. COMBUSTION AIR FLOW RATE SETTING = _____________
4. $\bar{V}_a$ = COMBUSTION AIR FLOW RATE

   $\bar{V}_a$ = __________ l/min AT S.T.P.
5. P_a = COMBUSTION CHAMBER PRESSURE

   P_a = __________ psig
6. ΔP_a = COMBUSTION CHAMBER PRESSURE DROP

   ΔP_a = __________ INCHES OF WATER
7. ρ_{ca} = COMBUSTION AIR DENSITY

   $$\rho_{ca} = (0.0808) \times \left(\frac{492}{460 + T_a}\right) \times \left(\frac{14.7 + P_a}{14.7}\right)$$

   ρ_{ca} = __________ lb/ft^3
8. V_a = CALCULATED COMBUSTION CHAMBER LINEAR VELOCITY

   $$V_a = (0.00125) \times \left(\frac{460 + T_a}{14.7 + P_a}\right) \times (\bar{V}_a)$$

   V_a = ___________ ft/sec

9. F_{ex} = FLUID BED EXPANSION FACTOR

F_{ex} = ____________

10. H_{fc} = DEPTH OF FLUIDIZED BED

H_{fc} = __________ft

11. VOL_{fb} = VOLUME OF FLUIDIZED BED

VOL_{fb} = __________ft^3

12. VOL_{cat} = VOLUME OF CATALYST IN FLUID BED

VOL_{cat} = _________ft^3

13. VOL_{void} = VOLUME OF FLUID BED NOT OCCUPIED BY CATALYST

VOL_{void} = __________ft^3

14. T_{ac} = AIR/CATALYST CONTACT TIME

$$T_{ac} = (56{,}871) \times \frac{(VOL_{void})}{(\overline{V}_a)} \times \frac{(14.7 + P_a)}{(460 + T_a)}$$

T_{ac} = __________SECONDS

15. T_a = TOTAL COMBUSTION TIME

T_a = __________MIN.

16. V_{ca} = VOLUME OF COMBUSTION AIR

$V_{ca} = \overline{V}_a \times T_a$

V_{ca} = ___________liters

E. SYSTEM PURGING

1. NITROGEN FLOW RATE SETTING = _____________
2. NITROGEN FLOW RATE = _________l/min (S.T.P.)
3. PURGING TIME = __________min
4. PURGE VOLUME = __________liters

S.T.P. = 32°F; 29.92 in Hg.

IX. ANALYTICAL DATA

A. H_2S ANALYSIS

1. V_t = VOLUME OF STANDARDS SODIUM THIOSULFATE TITRANT

V_{t_b} = ______ ml FOR BLANK

V_{t_s} = ______ ml FOR SAMPLE

2. N_t = NORMALITY OF STANDARD SODIUM THIOSULFATE TITRANT

N_t = ________ g-eq/liter

3. V_{SOLN} = TOTAL VOLUME OF SAMPLE + IODINE + ACID

V_{SOLN} = ______ ml

4. V_a = VOLUME OF ALIQUOT TITRATED

V_a = ______ ml

5. W_{H_2S} = WEIGHT OF SULFUR COLLECTED AS H_2S

$$W_{H_2S} = 16[(V_tN_t)_{BLANK} - (V_tN_t)_{SAMPLE}]\left(\frac{V_{SOLN}}{V_a}\right)$$

W_{H_2S} = ______MILLIGRAMS SULFUR

B. SO_2/SO_3 ANALYSIS

SULFUR COLLECTED AS ACID MIST AND SULFUR TRIOXIDE

1. V_t = VOLUME OF BARIUM PERCHLORATE TITRANT USED FOR SAMPLE

V_t= __________ml

2. V_{tb} = VOLUME OF BARIUM PERCHLORATE TITRANT USED FOR BLANK

V_{tb} = _________ml

3. N = NORMALITY OF BARIUM PERCHLORATE

N = __________ g - eq / liter

4. V_{soln} = TOTAL SOLUTION OF SULFURIC ACID, (FIRST IMPINGER + FILTER)

V_{soln} = ________ml

5. V_a = VOLUME OF SAMPLE ALIQUOT TITRATED

V_a = ________ml

6. $W_{H_2SO_4,SO_3}$ = WEIGHT OF SULFUR COLLECTED AS ACID MIST AND SULFUR TRIOXIDE

$$W_{H_2SO_4,SO_3} = 16\,\frac{(V_t - V_{tb})(\underline{N})(V_{soln})}{V_a}$$

$W_{H_2SO_4,SO_3}$ = __________MILLIGRAMS SULFUR

SULFUR COLLECTED AS SULFUR DIOXIDE

1. V_t = VOLUME OF BARIUM PERCHLORATE TITRANT USED FOR SAMPLE

2. V_t = __________ml

2. V_{tb} = VOLUME OF BARIUM PERCHLORATE TITRANT USED FOR BLANK

V_{tb} = __________ml

3. $\underline{N}$ = NORMALITY OF BARIUM PERCHLORATE TITRANT

$\underline{N}$ = __________ g - eq / liter

4. V_{soln} = TOTAL SOLUTION VOLUME OF SULFUR DIOXIDE (SECOND AND THIRD IMPINGERS)

V_{soln} = __________ml

5. V_a = VOLUME OF SAMPLE ALIQUOT TITRATED

V_a = __________ml

6. W_{SO_2} = WEIGHT OF SULFUR COLLECTED AS SULFUR DIOXIDE

$$W_{SO_2} = 16\,\frac{(V_t - V_{tb})(\underline{N})(V_{soln})}{V_a}$$

W_{SO_2} = __________MILLIGRAMS SULFUR

SULFUR COLLECTED AS SULFUR DIOXIDE

1. V_t = VOLUME OF BARIUM PERCHLORATE TITRANT USED FOR SAMPLE

   V_t = __________ml

2. V_{tb} = VOLUME OF BARIUM PERCHLORATE TITRANT USED FOR BLANK

   V_{tb} = __________ml

3. $\underline{N}$ = NORMALITY OF BARIUM PERCHLORATE TITRANT

   $\underline{N}$ = __________ g - eq / liter

4. V_{soln} = TOTAL SOLUTION VOLUME OF SULFUR DIOXIDE (SECOND AND THIRD IMPINGERS)

   V_{soln} = __________ml

5. V_a = VOLUME OF SAMPLE ALIQUOT TITRATED

   V_a = __________ml

6. W_{SO_2} = WEIGHT OF SULFUR COLLECTED AS SULFUR DIOXIDE

$$W_{SO_2} = 16 \frac{(V_t - V_{tb})(\underline{N})(V_{soln})}{V_a}$$

   W_{SO_2} = __________MILLIGRAMS SULFUR

X. SULFUR BALANCE AROUND STRIPPER

A. S_r = SULFUR CHARGED TO REACTOR

   S_r = __________MILLIGRAMS SULFUR

B. W_{H_2S} = WEIGHT OF SULFUR COLLECTED AS H_2S

   W_{H_2S} = __________MILLIGRAMS SULFUR

C. $W_{H_2SO_4,SO_3} + W_{SO_2}$ = FINAL WEIGHT OF SULFUR IN COKE

   $W_{H_2SO_4,SO_3} + W_{SO_2}$ = __________MILLIGRAMS SULFUR

D. $\%W_s$ = PERCENT OF FEED SULFUR ACCOUNTED FOR

$\%W_s$ = __________

XI. H_2O ANALYSIS - (TO BE DONE DURING STEAM STRIPPING)

A. TOTAL H_2O VOLUME IN IMPINGERS

V_{I_f} = FINAL VOLUME = __________ml (GRAMS)

V_{I_i} = INITIAL VOLUME = __________ml (GRAMS)

ΔV_i = VOLUME OF WATER COLLECTED IN IMPINGERS

$\Delta V_i = V_{I_f} - V_{I_i}$ = __________ml (GRAMS)

B. WEIGHT OF H_2O COLLECTED BY SILICA GEL

V_{sgf} = FINAL WEIGHT = __________GRAMS

V_{sgi} = INITIAL WEIGHT = __________GRAMS

ΔV_{sg} = WEIGHT OF WATER COLLECTED BY SILICA GEL

$\Delta V_{sg} = V_{sgf} - V_{sgi}$ = __________GRAMS

C. W_{H_2O} = TOTAL WEIGHT OF H_2O COLLECTED

$W_{H_2O} = \Delta V_i + \Delta V_{sg}$

W_{H_2O} = __________grams water

XII. RESULTS AND CONCLUSIONS

A. CATALYST CHARGED TO REACTOR = __________GRAMS

B. WATER USED IN STRIPPING = __________GRAMS

C. STEAM TO CATALYST RATIO = S/C

$$= \frac{\text{GRAMS } H_2O}{\text{100 GRAMS CATALYST}}$$

= ____________________

D. SULFUR CONTENT OF COKE INITIALLY = _______% BY WEIGHT

E. SULFUR CONTENT OF COKE AFTER STRIPPING = _______% BY WEIGHT

F. PERCENT SULFUR CONTENT OF COKE REDUCTION = ________%

G. DISCUSSION -

APPENDIX B

DETAILED COST ESTIMATES OF THE STEAM STRIPPING PROCESS

A detailed capital cost estimate was prepared for the typical case at a steam stripping rate of 6 kg H_2O/100 kg catalyst and 50,000 bpsd nominal size FCC unit. This estimate was obtained by cost estimating the major equipment necessary for the process, designated as purchased equipment cost in the following section. Fixed capital investment cost has been calculated by applying a fixed capital investment factor* of 4.8 to the purchased equipment cost (see Table B-13).

The capital investment cost for the worst case using the same size FCC unit (50,000 bpsd) has been calculated by assuming a scaling factor of 0.67 and applying it to the fixed capital investment for the typical case. The cost for the worst case is summarized in Table B-43. Capital investment cost estimates were made for both the typical and the worst case, and for 10,000 and 150,000 bpsd FCC unit nominal sizes at steam stripping rates of 6, 12, 40, and 100 kg H_2O/100 kg catalyst. The estimates were obtained by applying the scaling factor of 0.67 to the corresponding fixed capital investment cost figures determined for the 50,000 bpsd unit. The operating costs were prepared on an individual basis for each FCC unit size and steam stripping rate.

---

1 bpsd = 1.84×10^{-6} m^3/s.

---

*Peters, M. S., and K. D. Timmerhaus. Plant Design and Economics for Chemical Engineers, 2nd Edition. New York, McGraw-Hill Book Co., 1968. 850 pp.

For convenience, we have repeated the list of assumptions that were applied in the economic analyses in both the Phase I and Phase II evaluations. Detailed capital and operating cost estimates are presented in tabular form in the order of increasing FCC unit capacity and increasing steam stripping rates in Tables B-1 through B-60.

Since the original estimate in Phase I was made for a 45,000 bpsd catcracker we have revised it for the catcracker with 50,000 bpsd capacity.

B-1. ASSUMPTIONS USED FOR CAPITAL INVESTMENT COST ESTIMATE AND EQUIPMENT SIZE CALCULATIONS

- Catalyst-to-oil ratio (C/O) is 6 kg of catalyst per kg of total feed for the typical case, and 12 kg of catalyst per kg of total feed for the worst case
- Catalyst attrition rate is 3.33×10^{-4} kg per kg (0.1 lb of catalyst per barrel) of feed for the typical case and 6.66×10^{-4} kg per kg (0.2 lb of catalyst per barrel) of feed for the worst case
- Steam line pressure, 9.63×10^{5} Pa (125 psig)
- Sulfur content of coke before steam stripping, 1.5 wt %
- Sulfur content of coke after steam stripping, 0.243 wt %
- Stripper operating temperature, 811 K (1000°F)

---

1 barrel = 0.15899 m^3.
1 bpsd = 1.84×10^{-6} m^3.

- Stripper operating pressure, 3.43×10^{-5} Pa (35 psig)

- Velocity in stripper feed transfer lines, 12.2 m/s (40 ft/s)

- Vapor velocity in lines leaving the stripper, 30.5 m/s (100 ft/s)

- Velocity in stripper standpipe, 2.1 m/s (7 ft/s)

- Stripper bed density, 240 kg/m^3 (15 lb/cu ft)

- Catalyst bulk density, 801 kg/m^3 (50 lb/cu ft)

- Catalyst density in the standpipe, 561 kg/m^3 (35 lb/cu ft)

- Hydrogen sulfide produced in the steam stripper will be fed into existing Claus unit and no additional cost was assumed to be needed for expansion of this facility

- Velocity in stripper, 0.61 m/s (2 ft/s)

- Depth of fluidized bed in the stripper was assumed to be 3.05 m (10 ft) with the fluid bed occupying 50% of the total stripper volume

- Weight of FCC feed, 136.1 kg/barrel (300 lb/barrel)

- Fixed capital investment factor = 4.8

- Start-up cost - 10% F.C.I.*

---

1 barrel = 0.15899 m^3.

- Working capital - 10.5% F.C.I. *

- Interest on construction loan - construction period of 12 months; financed fixed capital at the rate of 8%/yr for average of half of construction period assumed

- Does not include sulfur recovery plant capital cost

- Base period - February 1973

- Scaling factor - 0.67

- CE plant cost index

1968	113.7
1969	119.0
1970	125.7
1971	132.3
1972	137.2
Feb. 1973	140.4

- Other assumptions used will be presented at the time of their use

---

* Fixed Capital Investment

B-2. DETERMINATION OF PURCHASED EQUIPMENT COST FOR 50,000 BPSD FCC UNIT, TYPICAL CASE

a. Catalyst Stripper

Catalyst Circulation Rate (CCR)

$$\text{CCR (lb/hr)} = \frac{\text{(catalyst to oil ratio) x (300 lb/bbl of oil) x (bbl/day of feed oil)}}{\text{(24 hr/day)}}$$

$$\text{CCR} = 75 \text{ x } (50{,}000) = 3.750 \text{ x } 10^6 \text{ lb/hr}$$
$$4.725 \text{ x } 10^2 \text{ kg/s}$$

Steam Stripping Rate (SR)

SR (lb/hr) = (lb of steam per lb of catalyst) x CCR

$$\text{SR} = 0.06 \text{ x } 3.75 \text{ x } 10^6 = 225{,}000 \text{ lb/hr}$$
$$28.3 \text{ kg/s}$$

Volumetric Flow of Steam (VF)

$$\text{VF} = \frac{\text{SR x 359 cu ft/lb mole x } \frac{(1000 + 460)}{460} \text{ x } \frac{14.7}{50}}{\text{18 lb/lb mole x 3600}}$$

$$\text{VF} = 5.17 \text{ x } 10^3 \text{ x } 225{,}000 = 1163 \text{ acfs}$$
$$32.9 \text{ m}^3\text{/s}$$

Cross-Sectional Area (A_L) and Diameter (D_L) of Feed Transfer Lines

$$A_L = \frac{\text{VF}}{\text{40 ft/s}} = 0.025 \text{ x } 1163 = 29.1 \text{ sq ft}$$
$$2.70 \text{ m}^2$$

$$D_L = \sqrt{\frac{4A_L}{\pi}} = 1.128 \text{ x } \sqrt{A_L} = 6.08 \text{ ft}$$
$$1.85 \text{ m}$$

---

1 barrel = 0.15899 m³.

Cross-Sectional Area (A_S) and Diameter (D_S) of the Stripper

$$A_S = \frac{VF}{2\ \text{ft/s}} = 0.5 \text{ x } 1163 = \begin{matrix} 582 \text{ sq ft} \\ 54.0\ \text{m}^2 \end{matrix}$$

$$D_S = 1.128 \text{ x } \sqrt{A_S} = \begin{matrix} 27.2 \text{ ft} \\ 8.29 \text{ m} \end{matrix}$$

Cross-Sectional Area (A_E) and Diameter (D_E of Lines Leaving the Stripper

$$A_E = \frac{VF}{100\ \text{ft/s}} = 0.01 \text{ x } 1163 = \begin{matrix} 11.63 \text{ sq ft} \\ 1.08\ \text{m}^2 \end{matrix}$$

$$D_E = 1.128 \text{ x } \sqrt{A_E} = \begin{matrix} 3.85 \text{ ft} \\ 1.17 \text{ m} \end{matrix}$$

Cross-Sectional Area (A_P) and Diameter (D_P) of Stripper Standpipe

$$A_P \text{ (sq ft)} = \frac{CCR}{3600 \text{ x (bed density) x (velocity)}}$$

$$A_P = \frac{3.750 \text{ x } 106}{3600 \text{ x } 35 \text{ x } 7} = \begin{matrix} 4.25 \text{ sq ft} \\ 0.39\ \text{m}^2 \end{matrix}$$

$$D_P = 1.128 \text{ x } \sqrt{A_P} = \begin{matrix} 2.33 \text{ ft} \\ 0.71 \text{ m} \end{matrix}$$

Catalyst Inventory (CI) in the Stripper

$$CI \text{ (lb)} = A_S \text{ x 10 ft x (stripper bed density)}$$

$$CI = 0.075 \text{ x } 582 = \begin{matrix} 87{,}300 \text{ lb} \\ 39{,}600 \text{ kg} \end{matrix}$$

The cost of the steam stripper was estimated based on weight of this equipment. The unit price for 5% Cr, 1/2% Mo steel, which was assumed to suit this application, was estimated at \$1.18/kg (53.6¢/lb). The weight of the steam stripper has been determined as 90,000 lb from which the cost was calculated:

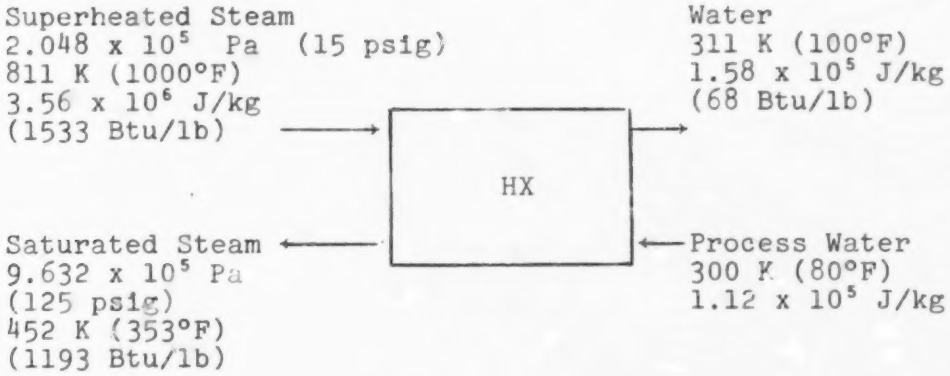
$$\$ = 0.536 \text{ x } 90{,}000 = \$48{,}240$$

b. Condenser

The area of the condenser was calculated according to the following assumptions.

- Stripping steam will be cooled from 811 K (1000°F) to 311 K (100°F) and condensed
- The cooling water at 300 K (80°F) will enter the condenser and will be evaporated to produce saturated steam at 452 K (353°F)
- The overall heat transfer coefficient was assumed to be 3973 W/m^2 K (700 Btu/ft^2 hr °F)
- The cost of heat exchanger was assumed at $7.75/sq ft (5% Cr, 1/2% Mo steel)

Calculation of Heat Transfer Area


Superheated Steam
2.048 x 10⁵ Pa (15 psig)
811 K (1000°F)
3.56 x 10⁶ J/kg
(1533 Btu/lb)
Water
311 K (100°F)
1.58 x 10⁵ J/kg
(68 Btu/lb)
HX
Saturated Steam
9.632 x 10⁵ Pa
(125 psig)
452 K (353°F)
(1193 Btu/lb)
Process Water
300 K (80°F)
1.12 x 10⁵ J/kg

Using the heat transfer equation

$$Q = UA \Delta T_m$$

Where $$\Delta T_m = \frac{(1000-353) - (100-80)}{\ln \frac{1000-353}{100-80}} = \begin{matrix}180°F \\ 355\ K\end{matrix}$$

$$Q = 225{,}000 \times (153-68) = \begin{matrix}329.6 \times 10^6\ Btu/hr \\ 9.657 \times 10^7\ W\end{matrix}$$

$$A = \frac{Q}{U\Delta T_m} = \frac{329.6 \times 10^6}{700 \times 180} = \begin{matrix}2616\ sq\ ft \\ 243\ m^2\end{matrix}$$

Cost \$ = 7.75 x 2616 = \$20,275

$$\text{Water requirement} = \frac{329.6 \times 10^6}{1193 - 48} \begin{matrix} = 287.9 \times 10^3\ lb/hr \\ = 0.363\ kg/s \\ = 575\ gpm\end{matrix}$$

The unit will produce 7.94 kg/s (63,000 lb/hr) of 9.63×10^5 Pa (125 psig) steam

The pump delivering this amount of water through the condenser and superheater operating at 9.63×10^5 Pa (125 psig) will have 37,300 W (50 HP)

c. Steam Superheater

It was assumed that the saturated steam from the condenser will be used as the feed for the superheater. The cost of the 2.63×10^7 W (90×10^6 Btu/hr) superheater was estimated at \$200,000.*
Natural gas was considered as the fuel.

Heat input required for the superheater was calculated as follows:

$$Q = 225{,}000\ lb/hr \times (1533 - 1193) = \begin{matrix}76.5 \times 10^6\ Btu/hr \\ 2.24 \times 10^7\ W\end{matrix}$$

---

*Private Communication with Struther-Wells Corporation

Superheater scale down:

$$\$ = 200{,}000\ \left(\frac{76.5}{90}\right)^{0.67} \quad = \$179{,}400$$

The amount of natural gas was approximated based on 15% heat loss and 3.722 x 10^7 J/m^3 (1000 Btu/cu ft) natural gas heating value:

$$\text{Natural gas} = \frac{1.15 \times 76.6 \times 10^6 \times 24}{1000} = \begin{array}{l} 2.11 \times 10^6 \text{ cu ft/day} \\ 0.692 \text{ m}^3\text{/s} \end{array}$$

d. Phase Separator

A 2.83 m^3 (750 gallon) stirred tank was assumed to be used for phase separation. The tank is made of 5% Cr, 1/2% Mo steel. The price of this tank was estimated at $6,600.

e. Acidifier

An epoxy-resin-lined, carbon steel, stirred, 2.83 m^3 (750 gallon) tank was assumed to suit this application. The cost of this tank was assumed to be $4,400. Acid sludge at 9.26 x 10^{-4} kg/s (7.35 lb/hr) is required to acidify the contents of the acidifier to pH 3. This amount was calculated based on the assumption that the sludge will contain 90% sulfuric acid.

f. Neutralizer

A 2.83 m^3 (750 gallon) carbon steel, stirred tank was assumed to suit this purpose. The cost of this tank was estimated to be $2,500. The amount of lime needed to neutralize the acid sludge was calculated to be 6.3 x 10^{-4} kg/s (5 lb/hr).

g. Clarifier

The mass flow rate through the clarifier was assumed to be

9.51 m^3/m^2 s (84 gal/sq ft hr). The area of the clarifier was determined as follows:

$$\frac{225{,}000 \text{ lb/hr}}{8.3 \text{ lb/gallon} \times 84 \text{ gallon/sq ft hr}} = \begin{array}{l}323 \text{ sq ft}\\ 30 \text{ m}^2\end{array}$$

From this the diameter of the tank was calculated to be 6.19 m (20.3 ft), or ∿20 ft. The depth of the clarifier was assumed to be 3.05 m (10 ft). The cost of this vessel was assumed to be $5,800.

h. Vacuum Filter

It was assumed that catalyst fines can be filtered by a vacuum filter operating at the load of 2.03×10^{-2} kg cake/m^2 s (15 lb cake/sq ft hr) with 70% moisture in the cake.

Assuming that 0.045 kg (0.1 lb) of catalyst per barrel of oil feed will be carried out from the steam stripper, the weight of filter cake and area of filter can be determined as follows:

$$\frac{0.1 \times 50{,}000}{24 \times 0.3} = \begin{array}{l}700 \text{ lb/hour}\\ 8.82 \times 10^{-2} \text{ kg/s}\end{array}$$

and

$$\frac{700}{15} = \begin{array}{l}46 \text{ sq ft}\\ 4.27 \text{ m}^2\end{array}$$

The cost of this equipment was assumed to be $14,500.

B-3. ASSUMPTIONS USED FOR OPERATING COST ESTIMATES

The operating cost estimates were individually calculated for each of the process operating conditions mentioned previously. The assumptions used to arrive at the final operating cost estimates are outlined as follows.

---

1 barrel = 0.15899 m^3.

- Catalyst loss was assumed at 3.33×10^{-4} kg/kg (0.1 lb of catalyst per barrel) of oil feed for the typical case and 6.66×10^{-4} kg/kg (0.2 lb of catalyst per barrel) of oil feed for the worst case

- The cost of catalyst was assumed at $600/ton

- Operating labor - 2 men per shift

- The cost of raw materials and utilities was assumed to change proportionally with the size of the FCC unit

- The cost of sulfuric acid was assumed at $40.8/ton (as 100% H_2SO_4)

- The cost of lime was assumed to be $19.50/ton

- Labor - $5.50/manhour

- Maintenance labor - 2% F.C.I.*

- Maintenance materials - 2% F.C.I.

- Process water - 7.93¢/m³ (30¢/1000 gal)

- Plant overhead - 80% total labor

- Taxes & insurance - [illegible]% F.C.I.

- G&A, sales, research - 6% F.C.I.

---

*F.C.I. = Fixed Capital Investment
1 barrel = 0.15899 m³.

- Depreciation - 10% F.C.I.
- Interest on working capital - 6% working capital
- Return on investment - 20%
- Value of steam - 0.047¢/10^6J (50¢/10^6 Btu)
- Stream factor - 0.9
- No credit for sulfur by-products was included in the cost estimates
- No correction was made to account for system retrofit expenses
- Necessary off-site facilities (such as Claus unit) are available
- Control laboratory labor - 10% of operating labor
- Operating materials - 10% of operating labor
- Fuel cost - 1.2¢/m^3 (34¢/1000 ft^3)
- Electricity - 0.278¢/10^6J (1¢/kw hr)

Table B-1. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 10,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

Fixed capital investment	\$360,100
Initial catalyst cost	4,400
Start-up cost	36,000
Working capital	37,800
Interest on construction loan	14,400
Total investment	\$452,700

---

1 bpsd = 1.84 x 10^6 m^3/s

Table B-2. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 10,000 pbsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $360,000
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	7,200
3	Control laboratory	19,300
4	Total labor	122,900
	Materials	
5	Raw and process - acid sludge	200
6	lime	100
7	catalyst replacement	99,000
8	Maintenance	7,200
9	Operating	9,600
10	Total materials	116,100
	Utilities	
11	Process water	11,000
12	Electricity	700
13	Fuel	31,700
14	Total utilities	43,000
15	Total direct operating costs (4, 10 & 14)	$282,400
	Indirect Operating Costs	
16	Plant overhead	98,300
17	Taxes and insurance	7,200
18	Plant cost (15, 16 & 17)	387,900
19	General & administrative, sales, research	21,600
20	Cash expenditures (18 & 19)	409,500
21	Depreciation	36,000
22	Interest on working capital	2,300
23	Total operating costs* (20, 21 & 22)	$447,800
24	Cost (cents/bbl)	13.84

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m³/s.

Table B-3. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 10,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

Fixed capital investment	$459,900
Initial catalyst cost	5,200
Start-up cost	45,900
Working capital	48,300
Interest on construction loan	18,400
Total investment	$577,700

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-4. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 10,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $459,900
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	9,200
3	Control laboratory	19,300
4	Total labor	124,900
	Materials	
5	Raw and process - acid sludge	300
6	lime	100
7	catalyst replacement	97,200
8	Maintenance	9,200
9	Operating	9,600
10	Total materials	116,400
	Utilities	
11	Process water	16,100
12	Electricity	900
13	Fuel	46,500
14	Total utilities	63,500
15	Total direct operating costs (4, 10 & 14)	$304,800
	Indirect Operating Costs	
16	Plant overhead	98,300
17	Taxes and insurance	9,200
18	Plant cost (15, 16 & 17)	413,900
19	General & administrative, sales, research	27,600
20	Cash expenditures (18 & 19)	441,500
21	Depreciation	46,000
22	Interest on working capital	2,900
23	Total operating costs* (20, 21 & 22)	490,400
24	Cost (cents/bbl)	15.14

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-5. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 10,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

Fixed capital investment	$731,700
Initial catalyst cost	10,500
Start-up cost	73,200
Working capital	76,800
Interest on construction loan	29,300
Total investment	$921,500

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-6. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 10,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $731,700
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	14,600
3	Control laboratory	19,300
4	Total labor	130,300
	Materials	
5	Raw and process - acid sludge	600
6	lime	200
7	catalyst replacement	97,200
8	Maintenance	14,600
9	Operating	7,700
10	Total materials	120,300
	Utilities	
11	Process water	32,200
12	Electricity	1,900
13	Fuel	93,000
14	Total utilities	
15	Total direct operating costs (4, 10 & 14)	$377,700
	Indirect Operating Costs	
16	Plant overhead	104,300
17	Taxes and insurance	14,600
18	Plant cost (15, 16 & 17)	496,600
19	General & administrative, sales, research	43,900
20	Cash expenditures (18 & 19)	540,500
21	Depreciation	73,200
22	Interest on working capital	4,600
23	Total operating costs* (20, 21 & 22)	$618,300
24	Cost (cents/bbl)	19.08

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-7. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 10,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

Fixed capital investment	$1,639,000
Initial catalyst cost	35,000
Start-up cost	163,900
Working capital	172,100
Interest on construction loan	65,600
Total investment	$2,075,600

---

1 bpsd = 1.84×10^6 m^3/s.

Table B-8. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 10,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $1,639,000
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	32,800
3	Control laboratory	19,300
4	Total labor	148,500
	Materials	
5	Raw and process - acid sludge	2,100
6	lime	800
7	catalyst replacement	97,200
8	Maintenance	32,800
9	Operating	9,600
10	Total materials	142,500
	Utilities	
11	Process water	119,100
12	Electricity	6,400
13	Fuel	309,500
14	Total utilities	435,000
15	Total direct operating costs (4, 10 & 14)	$ 726,000
	Indirect Operating Costs	
16	Plant overhead	118,800
17	Taxes and insurance	32,800
18	Plant cost (15, 16 & 17)	877,600
19	General & administrative, sales, research	98,300
20	Cash expenditures (18 & 19)	975,000
21	Depreciation	163,900
22	Interest on working capital	10,300
23	Total operating costs* (20, 21 & 22)	$1,150,100
24	Cost (cents/bbl)	35.50

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-9. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 10,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

Fixed capital investment	$3,029,000
Initial catalyst cost	87,600
Start-up cost	302,900
Working capital	318,000
Interest on construction loan	121,200
Total investment	$3,858,700

---

1 bpsd = 1.84×10^6 m^3/s.

Table B-10. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 10,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $3,029,000
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	60,600
3	Control laboratory	19,300
4	Total labor	176,300
	Materials	
5	Raw and process - acid sludge	5,600
6	lime	1,900
7	catalyst replacement	97,200
8	Maintenance	60,600
9	Operating	9,600
10	Total materials	174,900
	Utilities	
11	Process water	268,000
12	Electricity	16,000
13	Fuel	774,400
14	Total utilities	1,058,500
15	Total direct operating costs (4, 10 & 14)	$1,409,700
	Indirect Operating Costs	
16	Plant overhead	141,000
17	Taxes and insurance	60,000
18	Plant cost (15, 16 & 17)	1,611,300
19	General & administrative, sales, research	181,700
20	Cash expenditures (18 & 19)	1,793,000
21	Depreciation	302,900
22	Interest on working capital	19,100
23	Total operating costs* (20, 21 & 22)	$2,115,000
24	Cost (cents/bbl)	65.28

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-11. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 50,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

Fixed capital investment	$1,030,400
Initial catalyst cost	17,500
Start-up cost	103,000
Working capital	108,200
Interest on construction loan	41,200
Total investment	$1,300,300

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-12. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 50,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $1,030,400
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	20,600
3	Control laboratory	19,300
4	Total labor	136,300
	Materials	
5	Raw and process - acid sludge	1,100
6	lime	400
7	catalyst replacement	486,000
8	Maintenance	20,600
9	Operating	9,600
10	Total materials	517,700
	Utilities	
11	Process water	53,700
12	Electricity	3,200
13	Fuel	154,900
14	Total utilities	211,800
15	Total direct operating costs (4, 10 & 14)	$ 865,800
	Indirect Operating Costs	
16	Plant overhead	109,000
17	Taxes and insurance	20,600
18	Plant cost (15, 16 & 17)	995,400
19	General & administrative, sales, research	61,800
20	Cash expenditures (18 & 19)	1,057,200
21	Depreciation	103,000
22	Interest on working capital	6,500
23	Total operating costs* (20, 21 & 22)	$1,166,700
24	Cost (cents/bbl)	7.20

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-13. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 50,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

A.	Catalyst stripper	$ 48,250
B.	Condenser	20,270
C.	Steam superheater	179,400
D.	Phase separator	6,600
E.	Acidifier	4,400
F.	Neutralizer	2,500
G.	Clarifier	5,800
H.	Vacuum filter	14,500
	Total purchased equipment costs	$ 281,720
Fixed capital investment		$1,352,000
Initial catalyst cost		26,200
Start-up cost		135,200
Working capital		142,000
Interest on construction loan		54,100
	Total investment	$1,709,500

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-14. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 50,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $1,352,000
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	27,000
3	Control laboratory	19,300
4	Total labor	142,700
	Materials	
5	Raw and process - acid sludge	1,600
6	lime	600
7	catalyst replacement	486,000
8	Maintenance	27,000
9	Operating	9,600
10	Total materials	524,800
	Utilities	
11	Process water	80,600
12	Electricity	4,800
13	Fuel	232,400
14	Total utilities	317,800
15	Total direct operating costs (4, 10 & 14)	$ 985,300
	Indirect Operating Costs	
16	Plant overhead	114,200
17	Taxes and insurance	27,000
18	Plant cost (15, 16 & 17)	1,126,500
19	General & administrative, sales, research	81,100
20	Cash expenditures (18 & 19)	1,207,600
21	Depreciation	135,200
22	Interest on working capital	8,500
23	Total operating costs* (20, 21 & 22)	$1,351,300
24	Cost (cents/bbl)	8.34

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-15. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 50,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

Fixed capital investment	$2,151,000
Initial catalyst cost	52,400
Start-up cost	215,100
Working capital	225,900
Interest on construction loan	86,000
Total investment	$2,730,400

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-16. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 50,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $2,151,000
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	43,000
3	Control laboratory	19,300
4	Total labor	158,700
	Materials	
5	Raw and process - acid sludge	3,200
6	lime	1,100
7	catalyst replacement	486,000
8	Maintenance	43,000
9	Operating	9,600
10	Total materials	542,940
	Utilities	
11	Process water	161,200
12	Electricity	9,600
13	Fuel	464,900
14	Total utilities	635,700
15	Total direct operating costs (4, 10 & 14)	$1,337,300
	Indirect Operating Costs	
16	Plant overhead	127,000
17	Taxes and insurance	43,000
18	Plant cost (15, 16 & 17)	$1,507,300
19	General & administrative, sales, research	129,100
20	Cash expenditures (18 & 19)	1,636,400
21	Depreciation	215,500
22	Interest on working capital	13,600
23	Total operating costs* (20, 21 & 22)	$1,865,100
24	Cost (cents/bbl)	11.51

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-17. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 50,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

Fixed capital investment	$4,819,000
Initial catalyst cost	174,600
Start-up cost	481,900
Working capital	506,000
Interest on construction loan	192,800
Total investment	$6,174,300

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-18. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 50,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $4,819,000
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	96,400
3	Control laboratory	19,300
4	Total labor	212,100
	Materials	
5	Raw and process - acid sludge	10,600
6	lime	3,800
7	catalyst replacement	486,000
8	Maintenance	96,400
9	Operating	9,600
10	Total materials	606,400
	Utilities	
11	Process water	537,500
12	Electricity	32,100
13	Fuel	1,553,300
14	Total utilities	2,122,900
15	Total direct operating costs (4, 10 & 14)	$2,941,400
	Indirect Operating Costs	
16	Plant overhead	169,700
17	Taxes and insurance	96,400
18	Plant cost (15, 16 & 17)	3,207,500
19	General & administrative, sales, research	289,100
20	Cash expenditures (18 & 19)	3,496,600
21	Depreciation	481,900
22	Interest on working capital	30,400
23	Total operating costs* (20, 21 & 22)	$4,008,900
24	Cost (cents/bbl)	24.75

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-19. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 50,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

Fixed capital investment	$8,905,000
Initial catalyst cost	436,800
Start-up cost	890,500
Working capital	935,000
Interest on construction loan	356,200
Total investment	$11,523,500

---

1 bpsd = 1.84×10^6 m^3/s.

Table B-20. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 50,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $8,905,000
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	178,100
3	Control laboratory	19,300
4	Total labor	293,800
	Materials	
5	Raw and process - acid sludge	26,300
6	lime	9,500
7	catalyst replacement	486,000
8	Maintenance	178,100
9	Operating	9,600
10	Total materials	709,500
	Utilities	
11	Process water	1,343,700
12	Electricity	80,300
13	Fuel	3,888,600
14	Total utilities	5,312,600
15	Total direct operating costs (4, 10 & 14)	$6,315,900
	Indirect Operating Costs	
16	Plant overhead	235,000
17	Taxes and insurance	178,100
18	Plant cost (15, 16 & 17)	6,729,000
19	General & administrative, sales, research	534,300
20	Cash expenditures (18 & 19)	7,263,300
21	Depreciation	890,500
22	Interest on working capital	56,100
23	Total operating costs* (20, 21 & 22)	$8,209,900
24	Cost (cents/bbl)	50.68

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-21. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 150,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 4 kg H_2O/100 lb Catalyst

Fixed capital investment	\$2,209,000
Initial catalyst cost	65,300
Start-up cost	221,000
Working capital	232,000
Interest on construction loan	88,400
Total investment	\$2,816,600

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-22. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 150,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $2,209,900
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	44,200
3	Control laboratory	19,300
4	Total labor	159,900
	Materials	
5	Raw and process - acid sludge	3,300
6	lime	1,300
7	catalyst replacement	1,485,000
8	Maintenance	44,200
9	Operating	9,600
10	Total materials	1,543,400
	Utilities	
11	Process water	164,300
12	Electricity	10,700
13	Fuel	475,000
14	Total utilities	650,000
15	Total direct operating costs (4, 10 & 14)	$2,353,300
	Indirect Operating Costs	
16	Plant overhead	127,900
17	Taxes and insurance	44,200
18	Plant cost (15, 16 & 17)	2,525,400
19	General & administrative, sales, research	132,600
20	Cash expenditures (18 & 19)	2,658,000
21	Depreciation	221,000
22	Interest on working capital	13,900
23	Total operating costs* (20, 21 & 22)	$2,892,900
24	Cost (cents/bbl)	5.96

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-23. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 150,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

Fixed capital investment	\$2,822,600
Initial catalyst cost	78,700
Start-up cost	282,300
Working capital	296,400
Interest on construction loan	112,900
Total investment	\$3,592,900

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-24. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 150,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $2,822,600
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	56,400
3	Control laboratory	19,300
4	Total labor	172,100
	Materials	
5	Raw and process - acid sludge	4,700
6	lime	1,700
7	catalyst replacement	1,458,000
8	Maintenance	56,500
9	Operating	9,600
10	Total materials	1,530,500
	Utilities	
11	Process water	241,900
12	Electricity	14,500
13	Fuel	697,300
14	Total utilities	953,700
15	Total direct operating costs (4, 10 & 14)	$2,656,300
	Indirect Operating Costs	
16	Plant overhead	137,700
17	Taxes and insurance	56,500
18	Plant cost (15, 16 & 17)	2,850,500
19	General & administrative, sales, research	169,400
20	Cash expenditures (18 & 19)	3,019,900
21	Depreciation	282,300
22	Interest on working capital	17,800
23	Total operating costs* (20, 21 & 22)	$3,320,000
24	Cost (cents/bbl)	6.83

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-25. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 150,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

Fixed capital investment	$4,491,000
Initial catalyst cost	157,300
Start-up cost	449,100
Working capital	471,600
Interest on construction loan	179,600
Total investment	$5,748,600

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-26. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 150,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $4,491,000
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	89,800
3	Control laboratory	19,300
4	Total labor	205,500
	Materials	
5	Raw and process - acid sludge	9,500
6	lime	3,400
7	catalyst replacement	1,458,000
8	Maintenance	89,800
9	Operating	9,600
10	Total materials	1,570,300
	Utilities	
11	Process water	483,700
12	Electricity	28,900
13	Fuel	1,394,600
14	Total utilities	1,907,200
15	Total direct operating costs (4, 10 & 14)	$3,683,000
	Indirect Operating Costs	
16	Plant overhead	164,400
17	Taxes and insurance	89,800
18	Plant cost (15, 16 & 17)	3,937,200
19	General & administrative, sales, research	269,500
20	Cash expenditures (18 & 19)	4,206,700
21	Depreciation	449,100
22	Interest on working capital	28,300
23	Total operating costs* (20, 21 & 22)	$4,684,100
24	Cost (cents/bbl)	9.64

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-27. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 150,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

Fixed capital investment	$10,062,000
Initial catalyst cost	524,400
Start-up cost	1,006,200
Working capital	1,056,500
Interest on construction loan	402,500
Total investment	$13,051,600

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-28. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 150,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $10,062,000
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	201,200
3	Control laboratory	19,300
4	Total labor	316,900
	Materials	
5	Raw and process - acid sludge	31,600
6	lime	11,400
7	catalyst replacement	1,458,000
8	Maintenance	201,200
9	Operating	9,600
10	Total materials	1,711,800
	Utilities	
11	Process water	1,609,600
12	Electricity	96,400
13	Fuel	4,648,800
14	Total utilities	6,354,800
15	Total direct operating costs (4, 10 & 14)	$ 8,383,500
	Indirect Operating Costs	
16	Plant overhead	253,500
17	Taxes and insurance	201,200
18	Plant cost (15, 16 & 17)	8,838,200
19	General & administrative, sales, research	603,700
20	Cash expenditures (18 & 19)	9,441,900
21	Depreciation	1,006,200
22	Interest on working capital	63,400
23	Total operating costs* (20, 21 & 22)	$10,511,500
24	Cost (cents/bbl)	21.63

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-29. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 150,000 bpsd
Typical Stripping Operation
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

Fixed capital investment	$18,590,000
Initial catalyst cost	1,311,000
Start-up cost	1,859,000
Working capital	1,952,000
Interest on construction loan	743,600
Total investment	$24,455,600

---

1 bpsd = 1.84 x 10

Table B-30. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 150,000 bpsd at 90% Capacity
Typical Stripping Operation
Fixed Capital Investment: $18,590,000
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	371,800
3	Control laboratory	19,300
4	Total labor	487,500
	Materials	
5	Raw and process - acid sludge	79,100
6	lime	28,400
7	catalyst replacement	1,458,000
8	Maintenance	371,800
9	Operating	9,600
10	Total materials	1,946,900
	Utilities	
11	Process water	4,031,100
12	Electricity	241,000
13	Fuel	11,677,000
14	Total utilities	15,949,100
15	Total direct operating costs (4, 10 & 14)	$18,383,500
	Indirect Operating Costs	
16	Plant overhead	390,000
17	Taxes and insurance	371,800
18	Plant cost (15, 16 & 17)	19,145,300
19	General & administrative, sales, research	1,115,400
20	Cash expenditures (18 & 19)	20,260,700
21	Depreciation	1,859,000
22	Interest on working capital	117,100
23	Total operating costs* (20, 21 & 22)	$22,236,800
24	Cost (cents/bbl)	45.75

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m³/s.

Table B-31. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 10,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

Fixed capital investment	$572,900
Initial catalyst cost	8,700
Start-up cost	57,300
Working capital	60,200
Interest on construction loan	22,900
Total investment	$722,000

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-32. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 10,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $572,900
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	11,500
3	Control laboratory	19,300
4	Total labor	127,200
	Materials	
5	Raw and process - acid sludge	400
6	lime	200
7	catalyst replacement	198,000
8	Maintenance	11,500
9	Operating	9,600
10	Total materials	219,700
	Utilities	
11	Process water	22,000
12	Electricity	1,400
13	Fuel	63,400
14	Total utilities	86,800
15	Total direct operating costs (4, 10 & 14)	$433,700
	Indirect Operating Costs	
16	Plant overhead	101,800
17	Taxes and insurance	11,500
18	Plant cost (15, 16 & 17)	547,000
19	General & administrative, sales, research	34,400
20	Cash expenditures (18 & 19)	581,400
21	Depreciation	57,300
22	Interest on working capital	3,600
23	Total operating costs* (20, 21 & 22)	$642,300
24	Cost (cents/bbl)	19.82

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-33. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 10,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

Fixed capital investment	$731,700
Initial catalyst cost	10,500
Start-up cost	73,200
Working capital	76,800
Interest on construction loan	29,300
Total investment	$921,500

---

1 bpsd = 1.84×10^6 m^3/s.

Table B-34. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 10,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: \$731,700
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	\$ 96,400
2	Maintenance	14,600
3	Control laboratory	19,300
4	Total labor	130,300
	Materials	
5	Raw and process - acid sludge	600
6	lime	200
7	catalyst replacement	194,400
8	Maintenance	14,600
9	Operating	7,700
10	Total materials	217,500
	Utilities	
11	Process water	32,200
12	Electricity	1,900
13	Fuel	93,000
14	Total utilities	127,100
15	Total direct operating costs (4, 10 & 14)	\$474,900
	Indirect Operating Costs	
16	Plant overhead	104,300
17	Taxes and insurance	14,600
18	Plant cost (15, 16 & 17)	593,800
19	General & administrative, sales, research	43,900
20	Cash expenditures (18 & 19)	673,700
21	Depreciation	73,200
22	Interest on working capital	4,600
23	Total operating costs* (20, 21 & 22)	\$715,500
24	Cost (cents/bbl)	22.08

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-35. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 100,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

Fixed capital investment	$1,164,300
Initial catalyst cost	20,800
Start-up cost	116,400
Working capital	122,300
Interest on construction loan	46,600
Total investment	$1,470,400

---

1 bpsd = 1.84×10^6 m^3/s.

Table B-36. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 10,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $1,164,300
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	23,300
3	Control laboratory	19,300
4	Total labor	139,000
	Materials	
5	Raw and process - acid sludge	1,300
6	lime	500
7	catalyst replacement	194,400
8	Maintenance	23,300
9	Operating	9,600
10	Total materials	229,100
	Utilities	
11	Process water	64,400
12	Electricity	3,900
13	Fuel	186,200
14	Total utilities	254,500
15	Total direct operating costs (4, 10 & 14)	$622,600
	Indirect Operating Costs	
16	Plant overhead	111,200
17	Taxes and insurance	23,300
18	Plant cost (15, 16 & 17)	757,100
19	General & administrative, sales, research	69,900
20	Cash expenditures (18 & 19)	827,000
21	Depreciation	116,400
22	Interest on working capital	7,300
23	Total operating costs* (20, 21 & 22)	$950,700
24	Cost (cents/bbl)	29.34

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-37. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 10,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

Fixed capital investment	\$2,608,000
Initial catalyst cost	70,000
Start-up cost	260,800
Working capital	273,800
Interest on construction loan	104,300
Total investment	\$3,316,900

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-38. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 10,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $2,608,000
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	52,200
3	Control laboratory	19,300
4	Total labor	167,900
	Materials	
5	Raw and process - acid sludge	8,500
6	lime	1,500
7	catalyst replacement	194,400
8	Maintenance	52,200
9	Operating	9,600
10	Total materials	266,200
	Utilities	
11	Process water	214,400
12	Electricity	12,900
13	Fuel	619,100
14	Total utilities	846,400
15	Total direct operating costs (4, 10 & 14)	$1,280,500
	Indirect Operating Costs	
16	Plant overhead	134,400
17	Taxes and insurance	52,200
18	Plant cost (15, 16 & 17)	1,467,000
19	General & administrative, sales, research	156,500
20	Cash expenditures (18 & 19)	1,623,500
21	Depreciation	206,800
22	Interest on working capital	16,400
23	Total operating costs* (20, 21 & 22)	$1,846,700
24	Cost (cents/bbl)	57.00

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-39. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 10,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

Fixed capital investment	$4,819,000
Initial catalyst cost	174,600
Start-up cost	481,900
Working capital	506,000
Interest on construction loan	192,800
Total investment	$6,174,300

---

1 bpsd = 1.84×10^{6} m^3/s.

Table B-40. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 10,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $4,819,000
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	96,400
3	Control laboratory	19,300
4	Total labor	212,100
	Materials	
5	Raw and process - acid sludge	10,600
6	lime	3,800
7	catalyst replacement	194,400
8	Maintenance	96,400
9	Operating	9,600
10	Total materials	314,800
	Utilities	
11	Process water	536,100
12	Electricity	32,100
13	Fuel	1,553,300
14	Total utilities	2,121,500
15	Total direct operating costs (4, 10 & 14)	$2,648,400
	Indirect Operating Costs	
16	Plant overhead	169,700
17	Taxes and insurance	96,400
18	Plant cost (15, 16 & 17)	2,914,500
19	General & administrative, sales, research	289,100
20	Cash expenditures (18 & 19)	3,203,600
21	Depreciation	481,900
22	Interest on working capital	30,400
23	Total operating costs* (20, 21 & 22)	$3,715,900
24	Cost (cents/bbl)	114.7

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-41. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 50,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

Fixed capital investment	$1,639,400
Initial catalyst cost	35,000
Start-up cost	163,900
Working capital	172,100
Interest on construction loan	65,600
Total investment	$2,076,000

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-42. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 50,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $1,639,400
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	32,800
3	Control laboratory	19,300
4	Total labor	148,500
	Materials	
5	Raw and process - acid sludge	3,200
6	lime	1,100
7	catalyst replacement	972,000
8	Maintenance	32,800
9	Operating	9,600
10	Total materials	1,018,700
	Utilities	
11	Process water	161,200
12	Electricity	9,600
13	Fuel	464,800
14	Total utilities	635,600
15	Total direct operating costs (4, 10 & 14)	$1,802,800
	Indirect Operating Costs	
16	Plant overhead	118,800
17	Taxes and insurance	32,800
18	Plant cost (15, 16 & 17)	1,954,400
19	General & administrative, sales, research	98,400
20	Cash expenditures (18 & 19)	2,052,800
21	Depreciation	163,900
22	Interest on working capital	6,900
23	Total operating costs* (20, 21 & 22)	$2,223,600
24	Cost (cents/bbl)	13.73

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-43. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 50,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

Fixed capital investment	$2,151,000
Initial catalyst cost	52,400
Start-up cost	215,100
Working capital	225,900
Interest on construction loan	86,000
Total investment	$2,730,400

---

1 bpsd = 1.84×10^{6} m^3/s.

Table B-44. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 50,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $2,151,000
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

	Direct Operating Costs		
	Labor		
1	Operating		$ 96,400
2	Maintenance		43,000
3	Control laboratory		19,300
4	Total labor		158,700
	Materials		
5	Raw and process -	acid sludge	3,200
6		lime	1,100
7		catalyst replacement	972,000
8	Maintenance		43,000
9	Operating		9,600
10	Total materials		1,028,900
	Utilities		
11	Process water		161,200
12	Electricity		9,600
13	Fuel		464,900
14	Total utilities		635,700
15	Total direct operating costs (4, 10 & 14)		$1,823,300
	Indirect Operating Costs		
16	Plant overhead		127,000
17	Taxes and insurance		43,000
18	Plant cost (15, 16 & 17)		1,993,300
19	General & administrative, sales, research		129,100
20	Cash expenditures (18 & 19)		2,122,400
21	Depreciation		215,100
22	Interest on working capital		13,600
23	Total operating costs* (20, 21 & 22)		$2,351,100
24	Cost (cents/bbl)		14.51

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-45. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 50,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

Fixed capital investment	$3,422,600
Initial catalyst cost	104,900
Start-up cost	342,200
Working capital	359,400
Interest on construction loan	136,900
Total investment	$4,366,000

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-46. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 50,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $3,422,600
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	68,400
3	Control laboratory	19,300
4	Total labor	184,100
	Materials	
5	Raw and process - acid sludge	6,300
6	lime	2,270
7	catalyst replacement	972,000
8	Maintenance	68,400
9	Operating	19,300
10	Total materials	1,068,300
	Utilities	
11	Process water	322,500
12	Electricity	19,300
13	Fuel	929,800
14	Total utilities	1,271,600
15	Total direct operating costs (4, 10 & 14)	$2,524,000
	Indirect Operating Costs	
16	Plant overhead	147,300
17	Taxes and insurance	68,500
18	Plant cost (15, 16 & 17)	2,739,800
19	General & administrative, sales, research	205,400
20	Cash expenditures (18 & 19)	2,945,200
21	Depreciation	342,300
22	Interest on working capital	21,600
23	Total operating costs* (20, 21 & 22)	$3,309,100
24	Cost (cents/bbl)	20.43

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-47. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 50,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

Fixed capital investment	$7,668,000
Initial catalyst cost	349,200
Start-up cost	766,800
Working capital	805,100
Interest on construction loan	306,700
Total investment	$9,895,800

---

1 bpsd = 1.84×10^6 m^3/s.

Table B-48. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 50,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $7,668,000
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	153,400
3	Control laboratory	19,300
4	Total labor	269,100
	Materials	
5	Raw and process - acid sludge	21,100
6	lime	7,600
7	catalyst replacement	972,000
8	Maintenance	153,400
9	Operating	9,600
10	Total materials	1,163,700
	Utilities	
11	Process water	1,075,000
12	Electricity	64,200
13	Fuel	3,106,500
14	Total utilities	4,245,700
15	Total direct operating costs (4, 10 & 14)	$5,678,500
	Indirect Operating Costs	
16	Plant overhead	215,300
17	Taxes and insurance	153,400
18	Plant cost (15, 16 & 17)	6,047,200
19	General & administrative, sales, research	460,000
20	Cash expenditures (18 & 19)	6,507,200
21	Depreciation	766,800
22	Interest on working capital	48,300
23	Total operating costs* (20, 21 & 22)	$7,322,300
24	Cost (cents/bbl)	45.20

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-49. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 50,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

Fixed capital investment	$14,168,000
Initial catalyst cost	873,000
Start-up cost	1,416,800
Working capital	1,487,600
Interest on construction loan	566,700
Total investment	$18,512,100

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-50. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 50,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $14,168,000
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	283,400
3	Control laboratory	19,300
4	Total labor	399,100
	Materials	
5	Raw and process - acid sludge	52,900
6	lime	19,000
7	catalyst replacement	972,000
8	Maintenance	283,400
9	Operating	9,600
10	Total materials	1,336,900
	Utilities	
11	Process water	2,687,400
12	Electricity	160,600
13	Fuel	7,766,300
14	Total utilities	10,614,300
15	Total direct operating costs (4, 10 & 14)	$12,350,300
	Indirect Operating Costs	
16	Plant overhead	319,300
17	Taxes and insurance	283,400
18	Plant cost (15, 16 & 17)	12,953,000
19	General & administrative, sales, research	850,000
20	Cash expenditures (18 & 19)	13,803,000
21	Depreciation	1,416,800
22	Interest on working capital	89,300
23	Total operating costs* (20, 21 & 22)	$15,309,100
24	Cost (cents/bbl)	94.50

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-51. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 150,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

Fixed capital investment	$3,516,100
Initial catalyst cost	130,700
Start-up cost	351,600
Working capital	369,200
Interest on construction loan	140,600
Total investment	$4,508,200

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-52. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 150,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $3,516,100
Steam Stripping Rate: 4 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	70,300
3	Control laboratory	19,300
4	Total labor	186,000
	Materials	
5	Raw and process - acid sludge	6,700
6	lime	2,700
7	catalyst replacement	2,970,000
8	Maintenance	70,300
9	Operating	9,600
10	Total materials	3,059,300
	Utilities	
11	Process water	328,700
12	Electricity	21,300
13	Fuel	950,000
14	Total utilities	1,300,000
15	Total direct operating costs (4, 10 & 14)	$4,545,300
	Indirect Operating Costs	
16	Plant overhead	148,800
17	Taxes and insurance	70,300
18	Plant cost (15, 16 & 17)	4,764,400
19	General & administrative, sales, research	211,000
20	Cash expenditures (18 & 19)	4,975,400
21	Depreciation	351,600
22	Interest on working capital	22,200
23	Total operating costs* (20, 21 & 22)	$5,349,200
24	Cost (cents/bbl)	10.99

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-53. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 150,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

Fixed capital investment	$4,491,000
Initial catalyst cost	157,300
Start-up cost	449,100
Working capital	472,600
Interest on construction loan	179,600
Total investment	$5,748,600

---

1 bpsd = 1.84×10^{6} m^3/s.

Table B-54. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 150,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $4,491,000
Steam Stripping Rate: 6 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	89,800
3	Control laboratory	19,300
4	Total labor	205,500
	Materials	
5	Raw and process - acid sludge	9,500
6	lime	3,400
7	catalyst replacement	2,916,000
8	Maintenance	89,800
9	Operating	9,600
10	Total materials	3,028,300
	Utilities	
11	Process water	483,700
12	Electricity	28,900
13	Fuel	1,394,6000
14	Total utilities	1,907,200
15	Total direct operating costs (4, 10 & 14)	$5,141,000
	Indirect Operating Costs	
16	Plant overhead	164,400
17	Taxes and insurance	89,800
18	Plant cost (15, 16 & 17)	5,395,200
19	General & administrative, sales, research	269,500
20	Cash expenditures (18 & 19)	5,664,700
21	Depreciation	449,100
22	Interest on working capital	28,300
23	Total operating costs* (20, 21 & 22)	$6,142,100
24	Cost (cents/bbl)	12.64

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84×10^{-6} m^3/s.

Table B-55. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 150,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

Fixed capital investment	$7,145,400
Initial catalyst cost	314,600
Start-up cost	714,500
Working capital	750,300
Interest on construction loan	285,800
Total investment	$9,210,600

---

1 bpsd = 1.84 x 10^6 m^3/s.

Table B-56. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 150,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $7,145,400
Steam Stripping Rate: 12 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	142,900
3	Control laboratory	19,300
4	Total labor	258,600
	Materials	
5	Raw and process - acid sludge	19,000
6	lime	6,800
7	catalyst replacement	2,916,000
8	Maintenance	142,900
9	Operating	9,600
10	Total materials	3,094,300
	Utilities	
11	Process water	1,075,000
12	Electricity	57,900
13	Fuel	2,789,000
14	Total utilities	3,921,900
15	Total direct operating costs (4, 10 & 14)	$7,274,800
	Indirect Operating Costs	
16	Plant overhead	206,800
17	Taxes and insurance	142,900
18	Plant cost (15, 16 & 17)	7,624,500
19	General & administrative, sales, research	428,700
20	Cash expenditures (18 & 19)	8,053,200
21	Depreciation	714,500
22	Interest on working capital	45,000
23	Total operating costs* (20, 21 & 22)	$8,812,700
24	Cost (cents/bbl)	18.13

---

*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-57. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 150,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

Fixed capital investment	$16,009,000
Initial catalyst cost	1,048,800
Start-up cost	1,600,090
Working capital	1,680,900
Interest on construction loan	640,400
Total investment	$20,980,000

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1 bpsd = 1.84×10^{6} m^{3}/s.

Table B-58. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 150,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $16,009,000
Steam Stripping Rate: 40 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	320,200
3	Control laboratory	19,300
4	Total labor	435,900
	Materials	
5	Raw and process - acid sludge	63,300
6	lime	22,700
7	catalyst replacement	2,916,000
8	Maintenance	320,200
9	Operating	9,600
10	Total materials	3,331,800
	Utilities	
11	Process water	3,224,900
12	Electricity	192,800
13	Fuel	9,297,500
14	Total utilities	12,715,200
15	Total direct operating costs (4, 10 & 14)	$16,482,900
	Indirect Operating Costs	
16	Plant overhead	348,700
17	Taxes and insurance	320,200
18	Plant cost (15, 16 & 17)	17,151,800
19	General & administrative, sales, research	960,500
20	Cash expenditures (18 & 19)	18,112,300
21	Depreciation	1,600,900
22	Interest on working capital	100,900
23	Total operating costs* (20, 21 & 22)	$19,814,100
24	Cost (cents/bbl)	40.77

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*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 x 10^{-6} m^3/s.

Table B-59. SUMMARY OF CAPITAL INVESTMENT COSTS

FCC Unit Size: 150,000 bpsd
Worst Stripping Operation
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

Fixed capital investment	$29,579,000
Initial catalyst cost	2,622,000
Start-up cost	2,957,900
Working capital	3,105,800
Interest on construction loan	1,183,200
Total investment	$39,447,900

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1 bpsd = 1.84 x 10^6 m^3/s.

Table B-60. SUMMARY OF ANNUAL OPERATING COSTS

FCC Unit Size: 150,000 bpsd at 90% Capacity
Worst Stripping Operation
Fixed Capital Investment: $29,579,000
Steam Stripping Rate: 100 kg H_2O/100 kg Catalyst

	Direct Operating Costs	
	Labor	
1	Operating	$ 96,400
2	Maintenance	591,600
3	Control laboratory	19,300
4	Total labor	707,300
	Materials	
5	Raw and process - acid sludge	158,200
6	lime	56,700
7	catalyst replacement	2,916,000
8	Maintenance	591,600
9	Operating	9,600
10	Total materials	3,732,100
	Utilities	
11	Process water	8,062,200
12	Electricity	482,100
13	Fuel	23,243,800
14	Total utilities	31,788,100
15	Total direct operating costs (4, 10 & 14)	$36,227,500
	Indirect Operating Costs	
16	Plant overhead	565,800
17	Taxes and insurance	591,600
18	Plant cost (15, 16 & 17)	37,384,900
19	General & administrative, sales, research	1,774,700
20	Cash expenditures (18 & 19)	39,159,600
21	Depreciation	2,957,900
22	Interest on working capital	186,300
23	Total operating costs* (20, 21 & 22)	$42,303,800
24	Cost (cents/bbl)	87.04

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*Does not include by-product credit or recovery costs.
1 bpsd = 1.84 × 10^{-6} m^3/s.